PB86-138641

Proceedings: Symposium on Flue Gas Desulfurization (9th) Held at Cincinnati, Ohio June 1985. Volume 1. Sessions 1 through 5

Research Triangle Inst.
Research Triangle Park, NC

Prepared for

Environmental Protection Agency
Research Triangle Park, NC

Dec 85

U.S. Department of Commerce
National Technical Information Service
NTIS

PB86-138641

EPA/600/9-85/033a
December 1985

PROCEEDINGS: NINTH SYMPOSIUM ON FLUE GAS DESULFURIZATION
CINCINNATI, OHIO, JUNE 1985

Volume 1

Sessions 1 through 5

Compiled by

Franklin A. Ayer and Anna W. Wallace

Research Triangle Institute
P.O. Box 12194
Research Triangle Park, NC 27709

EPA Contract 68-02-3992, Task 1

Project Officers:

Julian W. Jones
U.S. Environmental Protection Agency
Air and Energy Engineering Research Laboratory
Research Triangle Park, NC 27711

Stuart M. Dalton
Electric Power Research Institute
3412 Hillview Avenue
Palo Alto, CA 94304

Prepared for:

U.S. Environmental Protection Agency
Office of Research and Development
Washington, DC 20460

Electric Power Research Institute
3412 Hillview Avenue
Palo Alto, CA 94304

REPRODUCED BY
NATIONAL TECHNICAL
INFORMATION SERVICE
U.S. DEPARTMENT OF COMMERCE
SPRINGFIELD, VA. 22161

TECHNICAL REPORT DATA

(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA/600/9-85/033a

2.

3. RECIPIENT'S ACCESSION NO. PB86 138641 /AS

4. TITLE AND SUBTITLE Ninth Symposium on Flue Gas Desulfurization, Cincinnati, Ohio, June 1985; Volume 1.. Sessions 1 through 5

5. REPORT DATE December 1985

6. PERFORMING ORGANIZATION CODE

7. AUTHOR(S) Franklin A. Ayer and Anna W. Wallace (compilers)

8. PERFORMING ORGANIZATION REPORT NO.

9. PERFORMING ORGANIZATION NAME AND ADDRESS

Research Triangle Institute
P.O. Box 12194
Research Triangle Park, North Carolina 27709

10. PROGRAM ELEMENT NO.

11. CONTRACT/GRANT NO. 68-02-3992, Task 1

12. SPONSORING AGENCY NAME AND ADDRESS

EPA, Office of Research and Development
Air and Energy Engineering Research Laboratory
Research Triangle Park, NC 27711

13. TYPE OF REPORT AND PERIOD COVERED Proceedings; 6/85

14. SPONSORING AGENCY CODE EPA/600/13

15. SUPPLEMENTARY NOTES AEERL project officer is Julian W. Jones, Mail Drop 61, 919/541-2489.

16. ABSTRACT The two-volume proceedings documents presentations at the EPA/EPRI Ninth Symposium on Flue Gas Desulfurization (FGD) in Cincinnati, OH, June 4-7, 1985. The meeting served as a forum for the exchange of technical and regulatory information and developments regarding systems and processes applicable to utility and industrial boilers. Addresses at the opening session: (1) compared regional declines of forests in Europe with those in North America and the possible role of airborne chemicals in these declines; (2) examined the current legislative/regulatory situation; and (3) gave results of a study of the economics of FGD systems in a variety of SO2 reduction strategies for coal-fired power plants. Subsequent technical sessions dealt with new and existing FGD technologies that enabled utility and industrial users of the technology, consultants, engineering firms, equipment manufacturers, process suppliers, government agencies, academia, and research and development firms to share their experiences.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group	
Pollution	Precipitation	Pollution Control	13B	04B
Flue Gases	Acidification	Stationary Sources	21B	07C
Desulfurization	Reliability	Acid Precipitation	07A, 07D	14D
Forest Trees	Economics	Dual Alkali Systems	06C, 02F	05C
Calcium Carbonates	Spray Drying		07B	13H
Organic Acids	Waste Disposal			

18. DISTRIBUTION STATEMENT Release to Public

19. SECURITY CLASS *(This Report)* Unclassified

20. SECURITY CLASS *(This page)* Unclassified

21. NO. OF PAGES 433

22. PRICE

EPA Form 2220-1 (9-73)

TABLE OF CONTENTS

VOLUME 1

Page

Session 1: OPENING SESSION 1
Michael A. Maxwell, Chairman

Comparison of Regional Declines of Forests in Europe and North America: The Possible Role of Airborne Chemicals . . . 2
Ellis B. Cowling

Address on Current Legislative/Regulatory Situation 21
Joseph A. Cannon

The Role of FGD in SO_2 Reduction Strategies for Coal-Fired Power Plants 22
E. S. Rubin*, M. Cushey, R. J. Marnicio,
C. N. Bloyd, and J. F. Skea

Session 2: COMMERCIAL STATUS OF FGD. 46
Stuart M. Dalton, Chairman

The Present and Future Status of FGD in the United States 47
Robert W. McIlvaine

Survey Report and Assessment of Utility Flue Gas Desulfurization System Installations in Europe. 63
William Ellison

Recent Developments in SO_2 and NO_x Abatement Technology in Japan. 75
Jumpei Ando

EPA's Stationary Source Control Technology Research Program 101
Carl R. Gerber

Session 3: LIMESTONE FGD/ORGANIC ACID ENHANCEMENT. 105
J. David Mobley, Chairman

Results of Using Organic Acid in San Miguel Electric's Flue Gas Desulfurization System 106
Jack M. Burke*, Robert Cmiel, and J. David Mobley

*Denotes speaker

Page

State-of-the-Art Design Applications on a Closed-Loop
FGD System. 121
Dennis Laslo* and Even Bakke

Dibasic Acid Test and Chemical Process Evaluation at
Petersburg Unit 3 FGD System. 136
D. Guetig, S. Ou, and C. P. Wedig*

Operating Results of Toyama Kyodo Electric Power's
Chiyoda Thoroughbred 121 Flue Gas Desulfurization System. 168
K. Wataya, A. Hori, N. Hashimoto, H. Koshizuka,
and D. D. Clasen*

Session 4, Part I: FGD RELIABILITY IMPROVEMENT 194
Dorothy A. Stewart, Chairwoman

The Sulfur Dioxide Removal System Improvement Program
at Seven Units of Louisville Gas and Electric Company 195
M. L. McInnis, D. A. Froelich, S. A. Bjorklun*,
and G. M. Graves

Improving the Reliability of Indianapolis Power & Light
Company's Petersburg 3 Limestone FGD System 202
J. David Colley*, Steve Wolsiffer, Dorothy Stewart,
Richard Rhudy, and David Balfour

Results of an FGD Process Troubleshooting Program at
Texas Utilities, Martin Lake Steam Electric Station 225
David Colley*, Don Mzyk, Mike Wadlington, and
Dorothy Stewart

Availability Analyses for Flue Gas Desulfurization
Improvement Decisions 247
Albert R. Cunningham* and Morton J. Smith

Session 4, Part II: CHEMISTRY/REAGENT PREPARATION. 260
Dorothy A. Stewart, Chairwoman

A Study of Sulfur-Nitrogen Compounds in Wet
Lime/Limestone FGD Systems. 261
James B. Jarvis*, Peter A. Nassos, and Dorothy A. Stewart

Thiosulfate as an Oxidation Inhibitor in Flue Gas
Desulfurization Processes: A Review of R&D Results 280
Gary T. Rochelle*, David R. Owens, John C. S. Chang,
and Theodore G. Brna

*Denotes speaker

Page

Lime Slaking Using Stirred Mills 302
N. N. Dharmarajan* and R. D. Forbus

Limestone Selection and Preparation for FGD 320
Dorothy Stewart*, J. David Colley,
O. W. (Buddy) Hargrove, and A. J. Jones

Session 5: MATERIALS OF CONSTRUCTION 338
Charles E. Dene, Chairman

Failure Analysis of FGD System Components 339
Gary D. Jones*, Peter F. Ellis II, Dennis M. Anliker,
and Dorothy A. Stewart

Performance of Duct and Stack Materials in Wet FGD Systems. . . . 362
H. S. Rosenberg*, G. H. Koch, C. W. Kistler, Jr.,
J. A. Beavers, M. L. Meadows, and D. A. Stewart

Leaning Brick Chimney Liners at Coal Fired Power Plants 387
E. R. Dille*, D. L. Krueger, and R. G. Rhudy

The Economics of Clad Metallic Fabrication Versus Solid
Alloy Metal Fabrication 399
E. H. Cloth* and R. Kreider

Laboratory Evaluation as a Technique for Predicting
the Behavior of Organic Coatings in FGD Systems 410
Malcolm L. White*, Henry Leidheiser, Jr., and G. H. Koch

VOLUME 2

Session 6: PANEL DISCUSSION ON RETROFITTING FGD SYSTEMS. 423
Frank T. Princiotta, Chairman

Session 7: DUAL ALKALI 429
Charles B. Sedman, Chairman

Development and Demonstration of a Limestone Dual Alkali
FGD Process at Central Illinois Public Service Company's
Newton Power Station. 430
Jerry L. Simpson* and James H. Wilhelm

Current Operating Experience and Operating Costs of
Double Alkali FGD at NIPSCO's R. M. Schahfer Generating
Station Unit 17 441
R. J. Biolchini*, W. L. Boward, Jr., and
B. S. Camponeschi

*Denotes speaker

Page

Current Progress at the FMC Limestone Double Alkali Demonstration Plant 455
Willard L. Boward, Jr.*, Robert J. Biolchini, and Beth A. Wrobel

Effects of Limestone Type and Grind on Dual Alkali System Performance 472
John C. S. Chang*, Theodore G. Brna, and Norman Kaplan

Session 8: EMERGING TECHNOLOGIES 501
Richard G. Rhudy, Chairman

Pilot Evaluation of Combined Particulate and SO_2 Removal Using a Fabric Filter System. 502
Richard G. Hooper*, Verle Bland, Franz G. Pohl, Michael McElroy, and Richard Rhudy

Investigation of Combined Particulate and SO_2 Using E-SOX 516
Leslie E. Sparks*, Norman Plaks, Geddes H. Ramsey, and Richard E. Valentine

Mitsui-BF Simultaneous SO_x and NO_x Removal System 531
Yoshiro Ito*, T. Fujimoto, and O. Nagaoka

Operating Experiences and Developments with Ljungstroem Gas-Gas Heaters in West German FGD Plants 544
David E. Clay

SOXAL^{tm} Process 559
K. N. Mani* and F. P. Chlanda

Session 9: SPRAY DRYER FGD 579
Theodore G. Brna, Chairman

Spray Dryer FGD Experience: Joy-Niro Installations 580
J. R. Donnelly*, S. Wilson, L. P. Matis, R. Eriksen, R. D. Emerson, and J. C. Fooks

Problem Solving and Results of Performance Test at United Power Association Stanton Station Dry FGD System. 602
Ghassem B. Manavizadeh*, Stefan Negrea, Gordon Westerlind, and Alan W. Ferguson

High-Sulfur Spray Dryer Evaluations 621
Russell F. Robards*, Robert W. Aldred, Thomas A. Burnett, Lynn R. Humphries, and Michael J. Widico

*Denotes speaker

Page

Recent Results from the EPRI 2-1/2 MW Spray Dryer Pilot Plant 643
Richard G. Rhudy and Gary M. Blythe*

Performance of a Spray Dryer/ESP Flue Gas Cleanup System During Testing at the Pittsburgh Energy Technology Center 670
James T. Yeh, Charles J. Drummond, James I. Joubert, Dennis M. Tulenko*, Robert R. Crynack, and Martin L. Hochhauser

Application of Electrostatic Enhancement of Fabric Filtration to Spray Dryer Byproduct 688
L. S. Hovis*, Bobby E. Daniel, Yang-Jen Chen, and R. P. Donovan

Comparison of Pilot Plant Data with Predictions from the U.S. EPA SPRAYMOD Computer Program. 702
Wayne T. Davis*, Gregory D. Reed, George P. Partridge, and Theodore G. Brna

Session 10: FGD BY-PRODUCT DISPOSAL/UTILIZATION. 715
Julian W. Jones, Chairman

Gypsum By-Product FGD System. 716
W. A. Liegois* and D. A. Wicks

By-Product Gypsum Production at a 2300 MW Power Plant 730
Don Myzk and Jan Zmuda
Paper presented by David Holstein

Evaluation of Engineering Properties and Wet Stacking Disposal of Widows Creek FGD Gypsum-Fly Ash Waste 748
John E. Garlanger*, Sal H. Magliente*, Thomas S. Ingra, and James L. Crowe

Solids Handling and Dewatering Optimization at Colorado Ute's Craig Station 767
O. W. (Buddy) Hargrove*, David R. Owens, Dorothy A. Stewart, and Jim Renner

EPRI Research on Management of Wastes From SO_2 Control Processes 785
Dean M. Golden* and Russell H. Boyd

Thermal Oxidation of Spray Dryer FGD Waste Product. 824
S. Bengtsson*, S. Ahman, T. Lillestolen, and G. Koudijs

*Denotes speaker

Page

UNPRESENTED PAPERS. 845

Reliability Problems and Solutions of FGD Systems 846
D. C. Agarwal

Four Corners 4 and 5 Waste Processing System: An Integrated and Unique Retrofit. 866
Alexander P. Simko, Richard R. Lunt, and Charles S. White

Corrosion Resistance Comparison of Duplex Stainless Steel with Type 317LM in Simulated Flue Gas Scrubber Environments. 878
Ronald E. Van Hoose and John B. Guernsey

Corrosion Resistance of a Cr-Ni-Mo-Cu-Fe Alloy in Scrubber Environments and Other Chemicals. 889
Terry A. DeBold, Douglas G. Frick, and Gerald A. Bauer

The Chemistry of Sodium Dry Sorbent Injection 902
A. W. Mueller and A. E. Winston

Techno-Economic Investigation Into the Hybrid Technology of FGD and PCC Applied to Large Boilers of Firing Canadian Coals. 922
R. L. Wang

Size Measurements in Aerosols Produced in Sulphur Dioxide Removal by Electron Beams 933
M. D. Carabine, P. G. Clay, and G. Sisniega

The NOXSO Process Development: An Update 948
J. L. Haslbeck and L. G. Neal

APPENDIX: Attendees. 976

PREFACE

These proceedings for the Ninth Symposium on Flue Gas Desulfurization (FGD) constitute the final report submitted to EPA's Air and Energy Engineering Research Laboratory (AEERL), Research Triangle Park, North Carolina, and to the Coal Combustion Systems Division, EPRI, Palo Alto, California. The symposium was conducted at the Netherland Plaza Hotel, in Cincinnati, Ohio, June 4 to 7, 1985.

Almost 600 representatives of electric utilities, state environmental agencies, equipment and process suppliers, coal and petroleum suppliers, architects, engineers, EPA, and other federal agencies, and research organizations attended the symposium.

The following persons were responsible for organizing this symposium:

- Julian W. Jones, Chemical Engineer, U.S. EPA, Air and Energy Engineering Research Laboratory, Gas Cleaning Division, Emissions Technology Branch, Research Triangle Park, North Carolina, was the EPA chairman.
- Stuart M. Dalton, Manager, Reliability and Nonrecovery Systems, EPRI, Coal Combustion Systems Division, Palo Alto, California, was the EPRI chairman.
- Franklin A. Ayer, Consultant, Research Triangle Institute, Research Triangle Park, North Carolina, was the symposium coordinator and compiler of the proceedings.

Session 6: PANEL DISCUSSION ON RETROFITTING FGD SYSTEMS

Frank T. Princiotta, Chairman
U.S. Environmental Protection Agency
Research Triangle Park, NC

TVA'S SO_2 CONTROL PROGRAM

by: Gerald A. Hollinden
Tennessee Valley Authority
Chattanooga, TN 37401

ABSTRACT

In the past few years, TVA has implemented various SO_2 emission control procedures on its coal-fired plants, such as coal cleaning, limestone-scrubbing flue gas desulfurization (FGD), and switching to low-sulfur coals. These emission reduction strategies, in combination with conservation and reduced power demand, improved heat rates, and additional nuclear generation, have greatly reduced the quantity of SO_2 emitted by TVA, with the 1983 SO_2 emissions being less than half the 1977 SO_2 emissions. Although great progress has been made, potential changes in SO_2 emission regulations to control acid rain could have profound effects on the use of FGD by TVA and other eastern utilities, e.g., increased use of FGD, upgrading of existing FGD systems, and the accelerated adoption of new technology. With a total FGD capacity of 2,500 MW already installed and test facilities for advanced FGD and combustion technology (AFBC) planned, TVA is continuing to assess controlling SO_2 emissions in the future. Even though limestone scrubbing is expected to be the primary method, at least through the rest of the century, innovations, such as the use of additives (adipic and dibasic acid), and new processes, such as limestone double alkali and dry FGD (spray dryer), should increase in importance and move into commercial use in the near future, particularly if the need for wider use of FGD occurs. AFBC, in which SO_2 removal is an integral function, is reaching an advanced stage of development and could serve to reduce the use of FGD at new facilities by the latter half of the 1990's.

Randall E. Rush
Southern Company Services
Birmingham, AL 35202

No abstract furnished.

A/E PERSPECTIVE OF RETROFITTING FGD SYSTEMS

by: Paul A. Ireland
Stearns Catalytic Corporation
Denver, CO 80217

ABSTRACT

From our perspective as engineers, we feel it is important to recognize the true cost of retrofitting an FGD system to an existing plant. We also feel that many acid rain related studies may be underestimating these costs. We also share the utility's concerns of minimizing outage time and maintaining existing plant services while the retrofit construction is underway. Often retrofit items can be overlooked such as having to take special precautions when dealing with existing asbestos insulation or having to hand dig foundations because of underground obstructions. In many instances, FGD process selection or design will be effected by the space available. Finally, if an FGD system retrofit is required for acid rain compliance, an opportunity may exist on some cases to minimize the overall cost by burning a cheaper higher sulfur coal.

THE CHANGING FGD MARKETPLACE

by: James R. Martin
Combustion Engineering Inc.
Birmingham, AL 35243

ABSTRACT

The demand in the United States for FGD systems at the beginning of this decade was concentrated primarily in the utility sector and consisted primarily of new systems in conjunction with utility capacity additions. During the last five years, we have seen that demand shrink at an average rate of 50% per year to a very low level. During the next five years, however, we can expect to see an increase in demand for FGD systems. In part, this new demand will be caused by utility capacity additions, but an increasing demand for emission control systems for refuse recovery, cogeneration, and hazardous waste incineration will have a major impact and FGD system rehabilitation and modifications will also be important.

This presentation will outline the forces that are reshaping the demand for gaseous emission control equipment and how we can expect the technology to adapt to that changing demand.

RETROFITTING FGD SYSTEMS

by: Howard Feibus
U.S. Department of Energy
Washington, DC 20545

ABSTRACT

In the event that legislation is enacted requiring control of SO_2 emissions from pre-NSPS coal-fired power plants in order to reduce acid rain precursor emissions, commercially available technology is essentially limited to coal cleaning, wet scrubbers, and spray dryers. DOE analyses indicate that commercial coal cleaning is of important but limited value and that the latter two would be prohibitively expensive. Developmental technologies that could be available in the late 1980's to mid-1990's time frame offer the potential of improved performance in terms of cost per unit weight of SO_2. These approaches include advanced coal cleaning processes, furnace sorbent injection, fluidized bed combustion, and duct injection FGD.

Jack F. Stewart
Babcock and Wilcox Co.
Barberton, OH 44203

No abstract furnished.

RETROFIT TECHNOLOGY OPTIONS FOR REDUCING SULFUR EMISSIONS

by: Stephen M. Katzberger
Sargent & Lundy Engineers
Chicago, IL 60603

ABSTRACT

A wide variety of technologies are available for reducing sulfur emissions from existing industrial and utility boilers. These technologies may reduce or capture sulfur: 1) prior to combustion through fuel switching, fuel blending, coal gasification, mechanical coal washing or chemical coal cleaning; 2) during combustion, as in furnace sorbent injection or fluidized bed boilers; or 3) after combustion, as in downstream sorbent injection, wet flue gas desulfurization (FGD) and spray drying. Each of these technologies has site specific advantages and disadvantages in terms of level of development, cost, retrofitability and sulfur reduction potential.

Each technology should be reviewed to determine site-specific impacts on the existing plant. Retrofits typically require unit outages to existing

plant equipment. Even fuel switching will require equipment modifications. Switching from high sulfur Illinois coal to low sulfur eastern Kentucky coal, for example, may require only minor modifications to ensure safe operation. However, switching to a low sulfur Powder River Basin coal can seriously impact coal handling, boiler performance, gas volume which affects both the fans and precipitator, ash resistivity and ash handling.

Unit outage time should be considered in evaluating retrofit options. Since a wet FGD system can be erected with the unit on line, only a 2-3 month outage would be needed to tie into existing ducts. A fluidized bed boiler retrofit could require a unit outage of 1½ to 2 years to modify a large part of the boiler.

The costs and impacts of alternative retrofit technologies should be thoroughly reviewed to select the best technology for each site. The sensitivity of the total evaluated cost of each alternative to the remaining life of the unit may show that high capital cost technologies, such as wet FGD, are not economically viable for older units without unit life extension. Low sulfur coal conversion, a lower capital cost-higher operating cost option may be selected for units with limited remaining lives or low capacity factors. Furnace sorbent injection might be selected for certain boiler designs and lower sulfur reduction requirements. Similarly, fluidized bed retrofits may be favored for units under 200 MW. In other cases, unit retirement may be the best choice.

A mixture of retrofit technologies is available with the best choice in each case depending on the age and size of the unit and the sulfur reduction required.

ISSUES IN UTILITY SO_2 CONTROL RETROFIT

by: Stuart M. Dalton
Electric Power Research Institute
Palo Alto, CA 94303

ABSTRACT

Cost, space, risk, energy, and waste disposal are the main issues facing utilities that are considering a retrofit FGD. Most debates have centered on FGD cost, and have normalized the cost in terms of dollars per ton of SO_2 control. Early estimates used in many political debates may have understated costs. EPRI estimates range from \$560 to \$1120 per ton SO_2 removed for conventional wet FGD and \$425 to \$935 for advanced concepts such as furnace sorbent injection and Chiyoda 121 scrubbing (levelized over 30 years).

Several factors cause costs to vary widely. Space constraints can cause costs to increase by two-thirds and will preclude FGD on a few installations. Utilities often are unwilling to accept high risk technologies since they cannot shut down if the process doesn't work. Electric energy used for FGD (1-3% of the plant output) is unavailable to meet the peak loads and be replaced by utilities at premium replacement rates. Waste disposal quantities for midwestern plants can double or even triple in volume compared to untreated plants.

These issues make it prudent for utilities to study the alternatives available to them very carefully before committing to a single strategy for their system.

Session 7: DUAL ALKALI

Charles B. Sedman, Chairman
U.S. Environmental Protection Agency
Research Triangle Park, NC

DEVELOPMENT AND DEMONSTRATION OF A LIMESTONE DUAL ALKALI FGD PROCESS AT CENTRAL ILLINOIS PUBLIC SERVICE COMPANY'S NEWTON POWER STATION

by: Jerry L. Simpson
Central Illinois Public
Service Company
607 East Adams Street
Springfield, IL 62701

James H. Wilhelm
Codan Associates
2394 Charros Road
Sandy, Utah 84092

ABSTRACT

The paper describes the development, implementation and demonstration of the CIPS/CODAN limestone dual alkali process at the 595 MW Newton Power Station of CIPS. The paper describes the process, the changes required to convert the original lime dual alkali process to the use of limestone, and the operating experience at Newton. Costs of development and retrofit are included along with operating cost savings, performance goals and achievements are included.

COMPANY AND STATION BACKGROUND

Central Illinois Public Service Company is an investor-owned utility which owns and operates five coal-fired generating stations with a combined capacity of about 2,800 megawatts. CIPS provides electricity to over 300,000 customers in 65 of Illinois' 102 counties. Illinois' coal reserves, accounting for 15% of the Nation's total and second in size only to those of Montana, are highly concentrated within our 20,000 square mile service area. Environmental constraints have required considerable efforts by CIPS to continue using high sulfur Illinois coal, especially for boilers subject to the New Source Performance Standards of the Clean Air Act, as is the case with CIPS' newest generating station, Newton.

Newton Unit 1 is a nominal 595 megawatt, pulverized coal-fired unit which began commercial operation during November, 1977. A dual alkali FGD system was placed in service during 1979 to control SO_2 emissions on Newton Unit 1. Since that time, many modifications and improvements were made to the scrubber which significantly improved its operation and availability. These changes, for which CIPS was granted a patent, have resulted in an exceptionally reliable, efficient FGD System. According to the Utility FGD Survey, conducted for EPRI by PEDCO, the Newton 1 FGD System availability is among the highest in the Nation of those using sulfur

Why Limestone Dual Alkali?

Excellent availability is just one of many beneficial attributes of our dual-alkali process. High SO_2 removal efficiency, improved waste handling characteristics and the elimination of slurry in the absorber vessels are other advantages of the dual alkali process as used at Newton. One might wonder why any attempt would be made to "improve" on such a system. Until now, to gain the benefit of a dual-alkali system, a utility was forced to use relatively expensive lime as a primary feedstock. Feedstocks, lime and soda ash, account for almost 1/2 of the annual Operation and Maintenance costs of the scrubber.

The desire to retain the advantages of the dual alkali system while reducing operating cost is responsible for CIPS' efforts toward developing a limestone-dual alkali process.

Limestone is the naturally occurring calcium carbonate from which quicklime or calcium oxide is produced. The product quicklime is subsequently more expensive than limestone due to the energy required to produce it. Lime is produced by calcining, or burning off, the carbon dioxide from limestone, according to this reaction:

$$CaCO_3 \xrightarrow{heat} CaO + CO_2$$

In CIPS' situation, quicklime is approximately three times more costly than limestone. Based on usage rates at Newton, the substitution of limestone for lime in the dual alkali process could result in potential operating savings in excess of two million dollars/year.

DEVELOPMENT OF LIMESTONE DUAL-ALKALI PROCESS

CIPS and CODAN began evaluation of the use of limestone in 1981. At that time there had been previous investigations of limestone dual alkali process undertaken by EPA and others, but there was no commercially viable process available for use at Newton. Past improvements in the Newton scrubber operation have been accomplished through the combined efforts of CIPS and CODAN Associates. These past successes, which lead to a patent for the Newton Lime Dual Alkali process, along with the potential economic benefits, were adequate incentive for us to begin development of the limestone dual alkali process.

An ongoing Task Force, made up of station personnel, and representatives from CIPS' Power Production, Environmental Affairs, and Construction Departments, which was instrumental in past scrubber improvements, provided input to the process development and testing undertaken by Jim Wilhelm of Codan Associates.

Early in the process development, the Task Force established goals to be achieved for limestone dual alkali to be feasible for use at Newton, including:

1) 90% Limestone utilization
2) Soda Ash consumption similar to that of the existing dual-alkali process (0.05 Moles soda ash/ Mole of SO_2 removed)
3) Maintain clarity of regenerated scrubbing liquor (less than 500 ppm suspended solids)
4) Similar performance to existing dual-alkali process -
    High SO_2 removal efficiency (approximately 0.6 lb/mmBtu)
    High availability (95% or greater)
    Good filtercake quality (over 60 wt% solids)
5) Flexibility to use lime or limestone

During nearly two years of lab-scale testing, conducted at the Salt Lake City laboratory of CODAN Associates, a process was developed which CIPS believed would merit full-scale testing. The resulting process is protected by patents issued to CIPS and licensed by CODAN Associates.

LIMESTONE DUAL-ALKALI CHEMISTRY

Careful consideration of the inherent differences between the lime and limestone reaction must be made before the new process could be successfully adapted to the Newton scrubber. The existing lime dual alkali process uses quicklime, slaked on-site, to regenerate sodium bisulfite formed when the scrubbing liquor absorbs SO_2 from the gas stream. The reaction is shown here:

$$2\ NaHSO_3 + Ca(OH)_2 \longrightarrow Na_2SO_3 + CaSO_3 + 2\ H_2O$$

The limestone dual alkali regeneration is as follows:

$$2NaHSO_3 + CaCO_3 \longrightarrow Na_2SO_3 + H_2O + CO_2\uparrow$$

Although it isn't obvious by comparing the reactions, the limestone reaction occurs at a much slower rate and at a lower pH. This necessitates longer residence time and therefore larger tank sizes. This would not be a problem when constructing a new FGD System, but did present problems to us because of the constraints of retrofitting an existing, operating scrubber.

SLUDGE BED REACTOR

One of the keys to making the limestone dual alkali process economically feasible at Newton was the development of the sludge bed reactor. This new reactor, designed by CODAN Associates, differs from the conventional stirred tank reactor in that it can allow sufficiently long residence times to be achieved within the sludge bed while maintaining normal liquor flow rates and minimizing tank size. Use of this device would permit use of tanks already existing at Newton by converting the existing thickeners to sludge bed reactors.

TESTING PARAMETERS

In addition to the effects of residence time, several other operating parameters were investigated during the lab-scale testing. These parameters included the effects of:

1) pH range
2) Active sodium level
3) Limestone grind
4) Presence of iron and magnesium
5) Sulfate concentration
6) Reactor configuration and residence time

As with most complex processes, there are parameters which operate in opposition to each other, requiring trade-offs to be made. For example, operation at lower pH levels increases the utilization of limestone but also reduces the scrubbing efficiency of the liquor. During the process development, these trade-offs were evaluated and operating parameters for the full-scale system selected.

PROCESS IMPLEMENTATION

With the basic process now developed, the Scrubber Task Force completed a preliminary engineering design and cost evaluation to implement the technology at Newton. This preliminary design was submitted to CIPS' management during the fall of 1983. Based on this preliminary design, capital costs to implement the process full-scale at Newton would be just over $1 million dollars. With a theoretical feedstock savings of about two million dollars per year, a successful project offered a payback period of less than one year.

Although the process had been tested only on a lab-scale, and was not commercially proven, CIPS management judged that the potential benefits outweighed the risks and approved implementation of the process at Newton. especially since the changes would still allow the flexibility for the system to be operated with lime.

PROCESS FLOW

The flow diagram to utilize the limestone dual alkali process at Newton is substantially different from that of the existing lime dual alkali process, yet retains the capability to return to the existing process.

Flue gas, after leaving the precipitator, is pushed by booster fans to the first absorbtion vessel, a co-current, open spray tower, previously used as a precooler for quenching and chloride removal. From the precooler, the gas flows upward through the scrubber absorber, a counter-current tray tower, and then through outlet ducts to the stack.

The liquor flow path starts with fresh, regenerated liquor entering the top of the scrubber absorber to maximize SO_2 removal. The liquor is continuosly recirculated and gravity flows to the precooler hopper. The precooler has its own pumps to provide recirculating flow to this spray tower, with a continuing bleed being sent back to the limestone reactors for regeneration.

Flexibility is present throughout the system, allowing continuing experimentation and optimization. The spent liquor, or bleed, is split between the primary sludge bed reactor (SBR) and the secondary sludge bed reactor. Limestone slurry can be introduced into the primary SBR. The primary sludge bed reactor is 50 feet in diameter with a capacity of 106,000 gallons. This reactor is equipped with a torque-actuated lifting rake. Liquor overflows and underflows from this SBR to the second SBR. Most of the reaction occurs in the primary SBR, causing most of the CO_2 to be evolved in this tank. This tank is covered to allow recovery of the CO_2 at a future date.

The existing three stirred tank reactors may be used to receive some of the bleed liquor, the underflow from the primary SBR, and limestone slurry, if desired. Operation in the limestone mode can continue with either the first SBR or stirred reactors out of service if necessary, and operation in the lime mode can be achieved using either reactor. The liquor from the stirred reactors overflows to the secondary SBR.

The secondary sludge bed reactor is a 100-foot-diameter sloped bottom tank with a capacity of 760,000 gallons. This reactor receives the liquor overflow from the primary SBR (and the stirred reactors when they are in service,) and is also capable of receiving scrubber bleed liquor. Solids settling in the second SBR are pumped to extractor filters for dewatering and disposal. The liquor overflows from the secondary SBR to the final clarifier.

The clarifier is identical in size and configuration to the secondary SBR and serves to remove suspended solids from the liquor. The solids which do settle are pumped to extractor filters for dewatering and disposal. Soda ash is added in the clarifier. Regenerated, clear liquor overflows the clarifier for reuse in the scrubbers.

MODIFICATIONS AT NEWTON

To accomplish this flow path and to compensate for the chemical differences required by the limestone regeneration, many changes were required at Newton. These changes can be generally categorized as follows:

1) Piping and pumping modifications
2) Linings
3) Adjustments and miscellaneous

Following the completion of detailed engineering during early 1984, work began procuring materials, fabricating components and making modifications to the scrubber. This work was completed over a six month period between February and August of 1984 with off-line tie-ins made during a six week scheduled maintenance outage.

The modifications made included the following:

1) Piping

   A) Rehabilitate regenerated liquor makeup to top scrubber spray header
   B) Tie precooler bleed into scrubber bleed
   C) Route total bleed to precooler thickener and "B" thickener
   D) Reroute limestone slurry to precooler thickener
   E) Route precooler overflow and underflow to "B" thickener
   F) Tie precooler pump discharge crossover tie to scrubber bleed valve
   G) Tie in bleed as soda ash makeup (new line)
   H) Modify emergency fresh water makeup to "B" thickener

2) Mechanical

   A) Modify bucket elevator for increased capacity and service factor
   B) Modify existing precooler thickener underflow pumps
   C) Rehabilitate scrubber recirculation pumps
   D) Upgrade screw conveyors
   E) Change size and speed of reaction tank agitators

3) Vessels/Tanks

   A) Make penetration for limestone slurry and bleed in precooler thickener dome
   B) Modify "A" and "B" thickeners, regenerated liquor tank and launders to allow changes in process flow
   C) Install "Y" pipes for limestone feed in precooler thickener
   D) Line both thickeners as required

4) Electrical

   A) Minor revisions and additions to existing heat tracing and piping insulation

5) Instrumentation

   A) Add pH probe in PC thickener outlet launder
   B) Increase speed of soda ash feeder to 3 TPH
   C) Add dual controls on ball mills to allow slaking/grinding
   D) Revise drawings to reflect instrumentation/process changes

6) Engineering

   A) Evaluate structural integrity of lime silos to hold limestone

PROJECT COSTS

The costs associated with the development, implementation and testing of the Limestone dual-alkali system at Newton are summarized as follows:

Process Studies, Development and Lab-Scale Testing	$ 204,000
Outside Engineering and Consulting	55,000
Purchased and Stock Materials for Modifications	269,000
Contractor Labor for Modifications	532,000
CIPS Payroll	55,000
Total Capital Expenditure (Not including AFUDC and Indirect Overhead)	$1,115,000

(The later addition of the hydroclone classification system added approximately $100,000 to this cost.)

SYSTEM STARTUP AND OPERATION

The modifications were completed ahead of schedule, with the first limestone introduced into the scrubber on August 2, 1984. Startup of the new process went well. Operation continued in the limestone mode for five days, when excessive wear problems with the thickener underflow pumps forced a switch back to quicklime. During the next few days of lime operation, structural modifications were made to the lime day bin to increase its load carrying capacity. A boiler tube leak forced the scrubber out of service and during this scrubber outage, the thickener underflow pumps were replaced with longer wearing low speed pumps.

Changes in the limestone grind, along with the new underflow pumps, prepared the system for a 2nd try at limestone operation, initiated on August 22, 1984. This run, however was shortened by another boiler tube leak, forcing the unit out of service.

Following rapid repairs to the boiler, the scrubber was started up in the limestone mode for the first time on August 25. (Previous limestone runs had been accomplished by converting from lime to limestone without removing the scrubber from service.) During this third run, which lasted 16 days, problems developed with the underflow pumps serving the first sludge bed reactor. These centrifugal pumps, which had not been changed for limestone operation proved to be inadequate for the task of transferring solids to the subsequent reactors. A positive displacement pump was installed during the test run and allowed continued operation until the scrubber was removed from service to make lining repairs to the stack.

The fourth through seventh test runs using the limestone process comprised another 29 days of limestone operation, bringing the total number of days of limestone operation during the initial test period to 51. All of these runs were terminated due to buildup of fine solids in the system.

Following the seventh test run, hydroclones were added to the ball mill system to reduce the amount of grit in the limestone. Operating and control procedures were also changed.

The system was once again switched to limestone on April 3, 1985. It ran through April 10, when a mechanical problem with the thickener rake drive caused shutdown of the thickener. During the test run the process operated well, with no buildup of fines. A second test run using limestone with the hydroclones in service was made between April 19 and May 2. This test run also provided very good results, with good regenerated liquor clarity and dry filtercake. The generating unit was removed from service for a scheduled maintenance outage on May 2, 1985. Continuous operation with limestone is planned following the scheduled annual outage which ends the middle of June.

A summary of the various test runs during the test period is included as Attachment No. 1.

TEST RESULTS

The following are the goals and the results achieved:

		OBJECTIVES	RESULTS ACHIEVED
1)	Limestone Utilization	90%	75 - 95%
2)	Soda Ash Consumption	0.05 Moles/Mole SO_2	0.05 Moles/Mole SO_2
3)	Filtercake Suspended Solids	60%	61 - 67%
4)	Regenerated Liquor Clarity	500 ppm	50 - 200 ppm
5)	Scrubbing Efficiency(approx.)	0.6 lb/mm BTU	0.6 lb/mm BTU
6)	Use of Lime or Limestone	Yes	Yes

The modifications have increased flexibility and improved operating characteristics in both lime and limestone modes of operation. The mode of operation can be easily and quickly switched from lime to limestone and back without removing the scrubber from service.

ATTACHMENT 1

SUMMARY OF NEWTON SCRUBBER OPERATION DURING LIMESTONE DUAL ALKALI TEST PERIOD AUGUST 2, 1984 - PRESENT

LIMESTONE DUAL ALKALI OPERATION

Run	Date	Duration (Days)	Reason for Test Run Conclusion
1	Aug. 2-7	5	Problems with thickener underflow pumps
2	Aug. 22-23	1	Unit outage - boiler tube leak
3	Aug.25-Sept.9	16	Scrubber outage - stack repairs
4	Sept. 19-26	7	Solids buildup - sulfate excursion
5	Oct. 18-28	10	Solids buildup
6	Oct.30-Nov.8	9	Solids buildup
7	Dec. 4-7	3	Solids buildup - sufficient (Data obtained pending hydroclone addition)
8	Apr. 3-10	8	Mechanical problem with thickener rake
9	Apr. 19-May 2	14	
		73	

LIME DUAL ALKALI OPERATION

Run	Date	Duration (Days)
1	Aug. 7-16	10
2	Aug. 20-22	2
3	Sept. 26-27	1
4	Oct. 10-18	8
5	Oct. 28-30	2
6	Nov. 8-17	8
7	Nov.25-Dec.4	9
8	Dec. 8-Apr. 2	115
9	Apr. 11-Apr. 18	8
		123

The testing indicates that the limestone dual-alkali process is a viable alternative which can offer the benefits of dual-alkali scrubbing with substantially reduced costs. Long term cost savings are projected to be up to two million dollars per year. These will be established during the next months and years of operation with limestone.

PROBLEMS EXPERIENCED

The problems experienced during the testing varied from mechanical problems occurring due to trying to make use of existing equipment not designed for its new purpose and inadequate limestone grind, to the production of fines solids and high oxidation. Many of the mechanical problems were resolved as the testing progressed by changing out existing equipment which was not properly suited for its new application. The limestone grind, judged to be inadequate and contributory to the buildup of solids in the system, was remedied with the addition of a hydroclone classification system. The buildup of fines is being controlled through adjustment of pressure drop through the scrubber, active sodium concentration and pH control.

The current use of limestone, initiated during April 1985, utilizes finer limestone grind and improved process control. This phase of testing resulted in continuous operation without buildup of fine solids and a more consistent, stable operation. While this next phase of testing continues, results are very encouraging. We have every reason to believe that the process will operate successfully in the long term at Newton, providing substantial savings over the life of the generating unit.

OTHER APPLICATIONS

Our experience with limestone dual alkali scrubbing has been only as it has been applied in a retrofit situation to an existing, top-performing dual-alkali system. Based on this experience, it is our opinion that this technology can be successfully and beneficially applied to new installations as well as retrofitted to existing lime/limestone systems to improve their availability and operation. Marketing rights to the process outside the CIPS system rest with CODAN Associates, and CIPS stands ready to work with interested parties in making our Newton facility available for testing and training. Our capabilities include providing engineering design and economic analysis for conversion of existing scrubbers to limestone dual alkali, as well as assistance with implementation and startup. CODAN and CIPS can provide the expertise gained from over 5 years of development and operation of the country's largest dual alkali scrubbers, and now the only operating full-scale limestone dual-alkali scrubber.

FUTURE PLANS

Present testing of the limestone dual alkali system will continue to determine long term cost savings and reliability. In addition, plans are being considered to recover the carbon dioxide that off-gases from the reactors. Preliminary investigations and testing of the relatively pure carbon dioxide gas indicate that it may be economically recovered and sold. Sales of the gas could pay for a majority, if not all, of the limestone used in the system.

CURRENT OPERATING EXPERIENCE AND OPERATING COSTS OF DOUBLE ALKALI FGD AT NIPSCO'S R. M. SCHAHFER GENERATING STATION UNIT 17

by: R. J. Biolchini,
W. L. Boward, and
B. S. Camponeschi
FMC Corporation
Schaumburg, Illinois 60195

ABSTRACT

This paper presents operating experience and 1984 operating costs of FMC's Double Alkali flue gas desulfurization system, for the 393MW, high sulfur coal-fired boiler, at NIPSCO's R. M. Schahfer Generating Station.

Test results have shown that the FGD system has met the EPA's 1979 Revised New Source Performance Standards on high sulfur coal and has demonstrated collection efficiencies well above 90%. Average availability during the first two years of operation has been 99.97 percent. FGD system operating costs for 1984 were on the order of 4 mils/kWh.

I. INTRODUCTION

Construction of Northern Indiana Public Service Company's Rollin M. Schahfer Generating Station Unit 17 began in 1981. This was the first unit burning high sulfur coal built under the revised New Source Performance Standard requiring 90% sulfur dioxide removal on a continuous 30-day rolling average. Unit 17 began commercial operation in April, 1983. There have been no SO_2 code violations or major load restrictions due to the FGD system since commercial operation began.

FMC provided NIPSCO with guarantees for SO_2 collection, particulate emissions, system pressure drop, reagent consumption, power consumption, waste product permeability, and system availability. All guarantees have been met or exceeded.

II. SYSTEM DESCRIPTION

NIPSCO's R. M. Schahfer Station is located on the Kankakee River near Wheatfield in northwestern Indiana. The station consists of Units 14 and 15, rated at about 500MW each, and Units 17 and 18, rated at 393MW each. Units 14 and 15 burn low sulfur compliance coal and do not require scrubbers. Units 17 and 18 are designed to burn 3.6% (dry) sulfur coal and utilize FMC's patented concentrated lime double alkali process for flue gas desulfurization. Unit 18 is under construction and is scheduled for a February, 1986 startup.

DESIGN CRITERIA

NIPSCO in conjunction with Sargent & Lundy established the major design criteria for the FGD system that are outlined in Table 1.

TABLE 1. FGD SYSTEM DESIGN CRITERIA

Design heat input to the boiler	3,967x 10^6 Btu/hr
Coal consumption based on design heat input and heating value of 10,500 Btu/hr	377,800 lb/hr
Coal sulfur range	2.5-3.8% (dry)
Flue gas flow at FGD system inlet	1,591,500 ACFM
Absorber inlet temperature	327°F
Absorbers	3 + 1 spare

Regeneration trains	Two trains (67% maximum load)
Flue gas reheat (inline steam)	30°F
SO_2 collection efficiency	90%
Particulate	No net addition at .03 lb/MM Btu

Other key design features included alloy construction of the absorbers and a fixation system for the FGD filter cake.

PROCESS DESCRIPTION

The NIPSCO system utilizes the conventional concentrated double alkali process furnished by FMC. Sulfur dioxide is absorbed according to the following reaction:

$$Na_2SO_3 + SO_2 + H_2O \longrightarrow 2NaHSO_3$$

Sodium bisulfite formed in the absorber is removed from the absorber loop as a bleed stream to a separate lime mixing tank where it is reacted with slaked lime to precipitate primarily calcium sulfite according to the following reaction:

$$2NaHSO_3 + Ca(OH)_2 \longrightarrow Na_2SO_3 + CaSO_3 \cdot \frac{1}{2}H_2O + 1\frac{1}{2}H_2O$$

The sodium sulfite that is regenerated by the above reaction is returned to the scrubbing process for reuse. The small amount of sodium that is not recoverable is replaced through the addition of soda ash to the scrubbing system. The pH of the sodium scrubbing solution is maintained at a level which buffers the scrubbing system and eliminates calcium scaling.

FACILITIES DESCRIPTION

The FGD system at NIPSCO is depicted schematically in Figure 1. Flue gas handling, including absorbers, dampers, and ductwork for Unit 17 and the future Unit 18 are independent of each other and under the control of boiler control room operators. The regeneration portion of the process, however, is integrated for Units 17 and 18 and under the control of FGD area operating personnel in an area several hundred feet away from the absorber area. The regeneration facility will ultimately contain four independent regeneration trains, each capable of accommodating two-thirds of the capacity of each boiler when operating on its maximum sulfur fuel. Under normal sulfur and load operating conditions, the regeneration area has effectively 100% spare capacity.

The flue gas leaves the boiler and enters the electrostatic precipitator. Immediately downstream of the precipitator are the induced draft fans which draw the flue gas through the boiler and pressurize the FGD

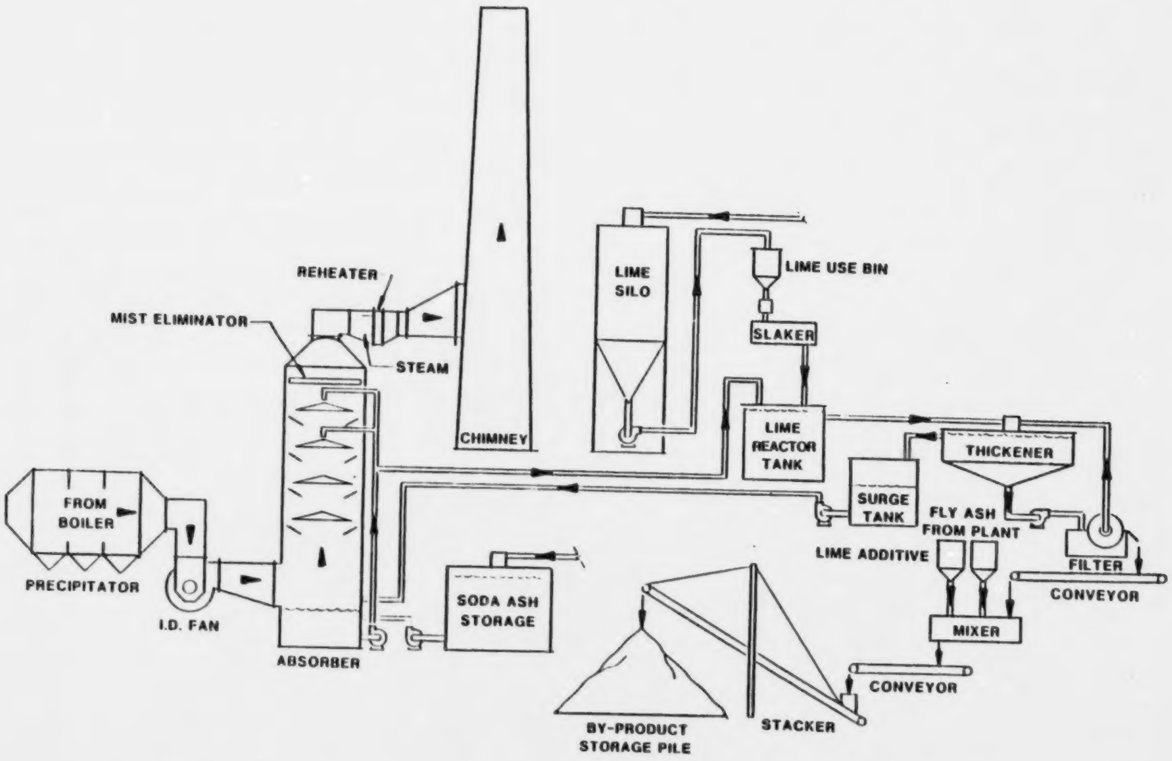

REHEATER
MIST ELIMINATOR
STEAM
CHIMNEY
LIME SILO
LIME USE BIN
SLAKER
LIME REACTOR TANK
THICKENER
FROM BOILER
PRECIPITATOR
I.D. FAN
ABSORBER
SODA ASH STORAGE
SURGE TANK
FLY ASH FROM PLANT
LIME ADDITIVE
FILTER
CONVEYOR
MIXER
CONVEYOR
BY-PRODUCT STORAGE PILE
STACKER

Figure 1. NIPSCO FGD process schematic

system. Flue gas enters the bottom of the 32-foot-diameter absorbers. These absorbers are constructed of 317LM stainless steel and consist of four disc and donut absorption stages. A concentrated solution of sodium sulfite is pumped from the sump in the base of each absorber to the top of the absorber at a liquid-to-gas ratio of approximately 10 gal/Macf. The scrubbing solution may be introduced at either of the top two stages. The solution cascades through the absorption stages while the gas flows countercurrent through the stages, to the top of the absorber. Two 100% capacity alloy (317LM) pumps per absorber continuously recirculate the solution and bleed off the spent solution to the regeneration area.

Installed in the upper level of each absorber is a mist eliminator constructed of Ryton high temperature plastic. From the mist eliminator, gas exits each absorber and passes through an in-line steam reheater, constructed of Inconel 625, which provides 30°F of reheat. From there the gas enters a common outlet duct of 316L stainless steel to the 500-foot-tall acid brick-lined chimney.

Proportionate to the amount of SO_2 collected, the absorber bleed stream is pumped to one of the two lime mixing tanks located in the regeneration building. There, slaked lime from the six slakers (2 spare) is added to the continuously agitated tanks to regenerate the sodium sulfite and form by-product solids. Each lime tank overflows by gravity to a 90-foot-diameter thickener which concentrates the calcium sulfite solids. Regenerated scrubbing solution overflows by gravity to a single regeneration surge tank, from which it is pumped on demand to the absorbers.

The thickener underflow is pumped by air-operated diaphragm pumps to four 33% capacity rotary drum vacuum filters. The filter cake is washed on the vacuum filters to recover additional sodium. Sodium remaining in the filter cake is replaced by adding soda ash (Na_2CO_3) in the regeneration return line from the surge tank.

The filter cake is transported by conveyor to one of the four mixers. The cake is mixed with flyash and lime to increase strength and reduce permeability.

The FGD by-product is conveyed to one of two radial stackers, serving both Units 17 and 18, and is discharged to a temporary disposal area. Then the material is loaded onto 50-ton off-road trucks for hauling to the on-site product disposal area located about one mile east of Unit 17. The material is placed and compacted to a planned maximum height of 67 feet. The approximately 200 acre disposal area will be covered with soil and seeded as each section is completed.

III. SYSTEM PERFORMANCE

FMC provided NIPSCO with guarantees for SO_2 removal, particulate emissions, FGD system availability, gas-side pressure drop, power consumption, chemical consumption, and waste product permeability. These guarantees have all been met or exceeded as illustrated by Table 2.

TABLE 2. CONTRACT GUARANTEE TEST SUMMARY

DATE	TEST RUN	SO_2 REMOVAL %	PARTICULATE EMISSIONS lb/MM Btu	SYSTEM ΔP in. w.g.	LIME STOICHIOMETRY	SODA ASH CONSUMPTION MOLE %	POWER CONSUMPTION kW
3/15/84	A-1	91.16	0.0106	--	--	--	--
3/16/84	A-2	91.96	0.0216	--	--	--	--
3/16/84	A-3	91.47	0.0142	--	--	--	--
3/17/84	A-4	93.06	0.0124	--	--	--	--
3/17/84	A-5	92.64	0.0109	--	--	--	--
7/23/84	B-1	92.87	0.0134	6.96	1.00034	1.940	1471
7/24/84	B-2	92.98	0.0075	7.21	1.00132	2.968	1563
7/24/84	B-3	91.95	0.0115	7.63	1.00032	1.853	1529
7/25/84	B-4	92.43	0.0086	7.02	1.00055	2.537	1563
7/25/84	B-5	92.43	0.0086	7.02	1.00055	2.537	1529
Average		92.30	0.0119	7.11	1.00072	2.335	1531
Contract Guarantee		90.0% Minimum	0.03 lb/MM Btu Maximum	9.4 in. w.g Maximum	1.01 Maximum	3.5% Maximum	2500kW Maximum

Guarantee testing was originally expected to occur in the summer and autumn of 1983, soon after the April startup. However, boiler problems began in July of 1983 and were not resolved until February of 1984. Guarantee testing was conducted from March through July of 1984, with the final results compiled in November, 1984.

The guarantee test periods as defined in the FMC/NIPSCO contract are:

Period 1	-	Start-up and Fine Tuning
Test A	-	SO_2 and Particulate Testing
Period 2	-	90 Day Availability Test
Test B	-	SO_2, Particulate, Consumables, and Waste Product Testing

SULFUR DIOXIDE COLLECTION

Contract guarantee sulfur dioxide testing was performed in Test Periods A and B in March and July of 1984, respectively. An SO_2 collection efficiency of 90% was guaranteed. An independent contractor did the stack testing, using EPA Methods 6 and 19. A total of five test runs were made over a three day interval during each test period. Only three of the four absorber stages were in service during these tests. The average SO_2 removal during Test A was 92.1% and for Test B it was 92.8%. The SO_2 removals for all ten stack tests ranged from 91.1% to 93.1%. These figures confirmed that Unit 17 SO_2 emissions were in compliance with the 1979 Revised New Source Performance Standards.

PARTICULATE

The particulate emission guarantee for the FGD system is 0.03 lb/MM Btu, under the condition that the absorber inlet particulate loading is 0.03 lb/MM Btu or less. Contract guarantee particulate and SO_2 testing was performed simultaneously during Test Periods A and B. EPA Methods 5 and 19 were used to determine particulate loading. The average loadings for Tests A and B were 0.0139, and 0.0101 lb/MM Btu, respectively.

PRESSURE DROP

A system pressure drop guarantee of 9.4 inches w.g. from the absorber inlet ductwork to the stack was made. Official contract testing yielded an average system pressure drop of 7.1 inches w.g. at full load. Absorber pressure drop is typically 4.5 to 5.0 inches w.g.

POWER CONSUMPTION

The FGD system power consumption, excluding gas-side pressure drop, was guaranteed not to exceed 2,500 kW. The average power consumption during Contract Test B was 1,531 kW, which is well below the guarantee.

The fan power required to overcome a system pressure drop of 7.1 inches w.g. is about 1,600 kW. Thus, the total FGD system power consumption, including fan power is roughly 3,100 kW. This corresponds to 0.8% of Unit 17's 393 MW generating capacity. Over 75% of the FGD system power usage is in the absorber area.

CHEMICAL CONSUMPTION

Soda ash consumption was guaranteed not to exceed 3.5 lb mole Na_2CO_3 per 100 lb mole SO_2 collected. Test B results yielded a consumption rate of 2.34 lb mole soda ash per 100 lb mole SO_2 collected. The soda ash consumption rate was determined based on filter cake analysis. During normal operation, all sulfur collected and sodium added to the FGD system exit in the filter cake. The sodium to sulfur ratio in the cake is therefore used to determine soda ash consumption rate.

Lime consumption rate was also determined by filter cake analysis. Lime consumption was guaranteed not to exceed 1.01 times stoichiometric requirements. Analysis of the filter cake for free lime showed a stoichiometry of 1.00072.

FILTER CAKE % SOLIDS

Filter cake discharged from the rotary vacuum filters typically has a 70% solids content. Figure 2 shows filter cake solids content for the first four months of 1985. The dry nature of this filter cake makes it easy to convey and handle. It may be passed through the pug mill mixer without fly ash addition, with no noticable deterioration in quality. Many similar filter cakes tend to liquify when subjected to shear. The low moisture content also contributes to a low soda ash consumption rate. The sodium lost in the entrained liquor of a 70% solids cake is roughly half of that for a 50% solids cake.

WASTE PRODUCT PERMEABILITY

The fixated filter cake/flyash/lime mixture was guaranteed to have a permeability below 5×10^{-6} cm/sec. Contract testing resulted in a mean permeability value of 0.2×10^{-6} cm/sec.

FGD SYSTEM AVAILABILITY

The Contract availability test was passed during the months of April-July of 1984. Table 3 shows an average system availability of 99.97% and SO_2 removal of 92.44% since start-up. The system downtime has been caused almost exclusively by materials handling problems. The pneumatic lime transfer system, lime slakers, thickener underflow pumps, and rotary vacuum filters have all contributed to lost availability. Reliability of the absorbers has been excellent due to clear liquor scrubbing.

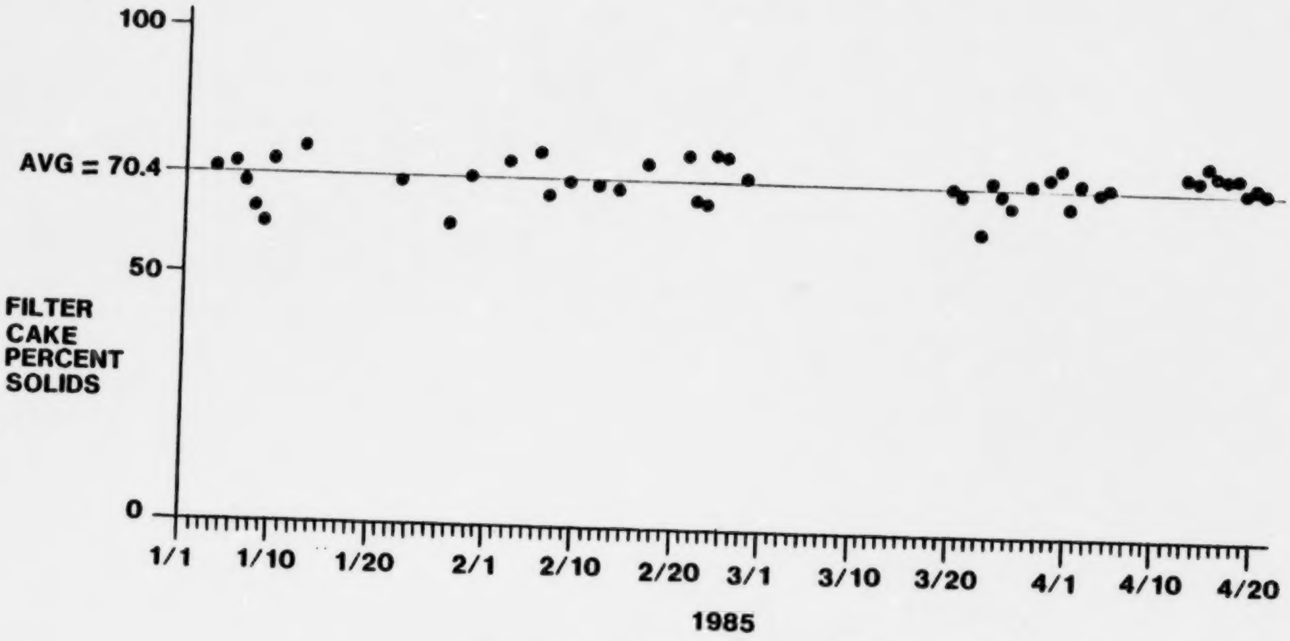


FIGURE 2. FILTER CAKE PERCENT SOLIDS

As a point of reference for the FGD system availability, it should be noted that Unit 17 overall availability for 1984 was 85.8%. Unit 17 had a capacity factor of 63.7% for 1984, which was second highest in the NIPSCO system. Thus, NIPSCO does not consider the FGD system a hinderance to Unit 17's usefulness.

TABLE 3. MONTHLY AVAILABILITY AND
SO_2 COLLECTION OF FGD SYSTEM

	Availability	Average SO_2 Removal
July, 1983	100.0%	91.2
August	100.0%	92.1
September	100.0%	93.2
October	100.0%	N/O
November	100.0%	92.3
December	100.0%	92.4
January, 1984	100.0%	N/O
February	99.7%	91.9
March	99.9%	93.0
April	99.9%	93.0
May	100.0%	93.4
June	100.0%	93.2
July	100.0%	92.2
August	100.0%	92.4
September	100.0%	92.2
October	99.9%	92.8
November	100.0%	93.1
December	100.0%	92.8
January 1985	100.0%	92.3
February	100.0%	91.3
March	100.0%	91.8
April	100.0%	92.2
Average	99.97%	92.44

N/O - No Boiler Operation

OPERATING COSTS

FGD system operating costs have been calculated for 1984 (1), and are shown in Table 4. All costs are based on the actual power generated at Schahfer Unit 17. The chemical costs are based on shipments received at the plant. Operating labor includes one supervisor and five operators per shift. Maintenance costs have been estimated based on fourth quarter data. Disposal costs are included in the operating and maintenance figures.

Power cost is based on an FGD power usage of 0.8% of the total plant generating capacity and a charge rate of 45 mils/kWh. This includes the estimated fan power to overcome the FGD system pressure drop, although there is not a separate absorber booster fan. Reheater steam usage costs have not been included.

TABLE 4. 1984 FGD SYSTEM OPERATING COSTS

	Mils/kWh
Soda Ash	0.43
Lime	1.88
Operating Labor	0.57
Mainterance	0.68
Power°	0.36
Total	3.92

° $$\frac{393 \text{ MW (Gross)} \times 0.637 \times 0.008 \times 45 \text{ mils/kWh}}{393 \text{ MW (Gross)} \times 0.637} = \frac{0.36 \text{ mils}}{\text{kWh}}$$

where 0.637 = Unit 17 capacity factor

0.008 = FGD system fractional power consumption

IV. EQUIPMENT OPERATING PROBLEMS AND SOLUTIONS

Several FGD system operating problems have arisen and been solved since the 1983 startup. Some of these problems were reported on at the 1983 Symposium in New Orleans (2) and other more recent conferences (3) (1). Two of the most current modifications will be covered in this section.

DUCTWORK LINING

A small section of 316L (2.75% Molybdenum minimum) ductwork at the outlet of each absorber has been lined with 317LM plate. This section of ductwork is located between the in-line reheater and the outlet isolation damper, as shown on Figure 3. It represents about 7% of the total outlet ductwork area.

The original 316L material in the converging side walls of this section experienced varying degrees of pitting. The floor and ceiling were relatively corrosion free. It is felt that the pitting was aggravated by the impingement of entrained scrubbing liquor droplets on the converging sidewalls. The scrubbing liquor typically has a chloride concentration of 7,000 ppm. The dissolved salts in these droplets were subsequently baked

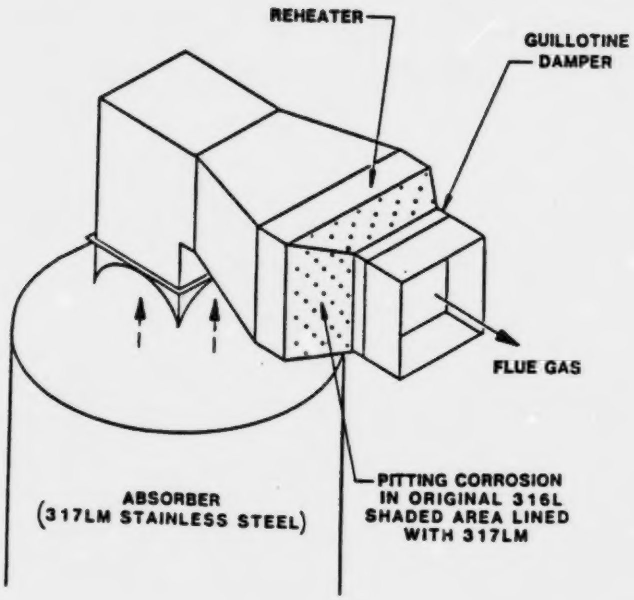

REHEATER
GUILLOTINE DAMPER
FLUE GAS
ABSORBER (317LM STAINLESS STEEL)
PITTING CORROSION IN ORIGINAL 316L SHADED AREA LINED WITH 317LM

Figure 3. Isometric View of Absorber Outlet Ductwork Showing Lined Section

onto the metal by the in-line reheater. This acidic, high chloride environment resulted in pitting. The pitting was most likely accelerated by frequent cycling of absorbers in and out of service during startup. This led to alternating wet and dry conditions on the 316L material. Also, the reheater was kept in service during short scrubber shutdowns. With superheated steam temperatures as high as 700°F, this led to high ductwork temperatures in the effected area. This accelerated the pitting as proven by the fact that pitting was worst close to the reheater.

All four converging ductwork sections were lined with 3/16 inch 317LM plate by early April, 1985. There has already been some pitting found in the lined ductwork. However, a sample of the pitted lining was analyzed and found to be 316 L which was inadvertently installed. The amount of pitted metal which is actually 317 LM is unclear at this time. The service life of 317 LM in this corrosive environment is currently under evaluation. Preliminary test results indicate that full service life with minimal maintenance is attainable with 317 LM.

Despite this ductwork lining, the overall choice of 317LM for the absorbers and 316L for the outlet plenum stands as a cost effective one. There has been virtually no evidence of corrosion in the absorbers themselves, and only minor pitting in the outlet plenum.

CAKE MIXER DISCHARGE CHUTES

Each of the four filter cake/flyash mixers has a 12 foot long discharge chute to an inclined belt conveyor. These chutes were continually plugging with either unmixed filter cake or the fixated waste product. After several unsuccessful attempts to alleviate the plugging with minor changes, the entire chute was redesigned.

Figure 4, shows the original and revised chute. The original chutes were only a foot wide. They were lined on the two narrow sides with polypropylene. A rotary plugged chute switch was installed near the bottom of the chute. These chutes would routinely plug within an hour after mixer startup. The solids leaving the mixer would quickly begin sticking to the chute walls, including the plastic liners. The entire cross section of the chute would soon bridge over. This usually occurred above the plugged chute switch, so the first indication of pluggage was waste product coming out of the mixer inspection door. The easiest way to clear the pluggage was usually to flush it away with water. Ironically, once the chute walls were wet, they tended to accumulate solids and plug even faster.

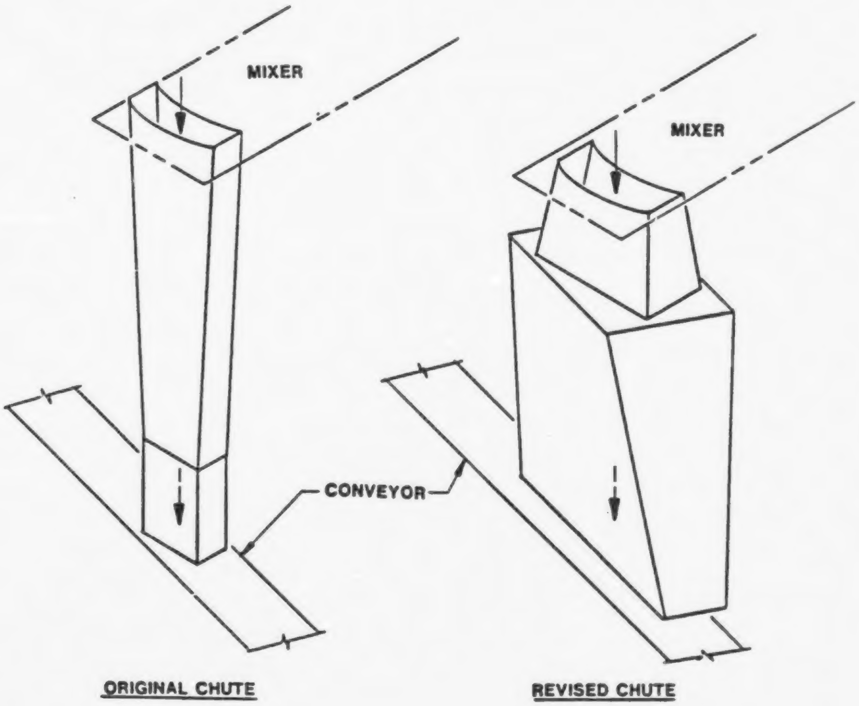


Figure 4. Isometric View of Original and Revised Mixer Discharge Chute

The first attempt to solve this problem was to install vibrators on the chute walls. This decreased the frequency of pluggage only slightly. Next, free-hanging rubber curtains were installed along the inside of the chute walls. These seemed to work quite well, until solids got behind the rubber. Another revision was an open wall chute consisting of sheet rubber walls within an angle iron framework. In each case, a lot of operator attention was still required to keep these chutes open.

It was decided to revise the entire chute, as shown on Figure 4. The major improvements are listed below:

- Increase size of mixer tub discharge opening.
- Make chute in two sections to accommodate angle between mixer and conveyor.
- Make chute walls diverging wherever possible.
- Line chute with Teflon.

The revised chutes have been in service since early March with no complete blockage to date.

V. SUMMARY

The first two years of FGD operations at Schahfer Unit 17 have been characterized by routine SO_2 compliance and high availability. This is due in part to the sparing philosophy, process design, and most importantly, the commitment of both NIPSCO and FMC to make the FGD system work. Problems have usually been solved in a timely manner with minimum impact on availability. We look forward to continued success at Unit 17, and a repeat performance for FGD at the impending start-up of Schahfer Unit 18.

The work described in this paper was not funded by the U. S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

REFERENCES

1. Steiner, T. P., "Dual Alklai Experience at Northern Indiana Public Service Company,"Power Magazine's Acid Rain Conference, March, 1985.

2. Glancy, D. L., Grant, R. J., Legatski, L. K., Wilhelm, J. H., Wrobel, B. A., "Utility Double Alkali Operating Experience," EPA/EPRI Symposium on Flue Gas Desulfurization, New Orleans, LA, November, 1983.

3. Coons, J. D., Rorke, K. C., Rosene, C. A., "Double Alkali Operating Experience at the NIPSCO R. M. Schahfer Generating Station," A.S.M.E's Industrial Power Conference, Toronto, Ontario, September-October, 1984.

CURRENT PROGRESS AT THE FMC LIMESTONE DOUBLE ALKALI DEMONSTRATION PLANT

By: Willard L. Boward, Jr.
Manager, Process Engineering and
Robert J. Biolchini
Senior Process Engineer
FMC Corporation
Air Quality Control Operation
Schaumburg, Illinois 60195

Beth A. Wrobel
Mechanical Engineer
Northern Indiana Public Service Company
Hammond, Indiana 46302

ABSTRACT

The Limestone Double Alkali (LSDA) flue gas desulfurization process combines the reliability and efficiency of sodium scrubbing with the economical use of limestone as the main reagent. Studies sponsored by EPA and EPRI have shown LSDA to be economically desireable for full scale plants when evaluated against other types of wet scrubbers. An LSDA demonstration plant has been in operation since September, 1984, at Northern Indiana Public Service Company's R. M. Schahfer Station. The demonstration plant treats a flue gas slip stream from Schahfer Unit 17 which burns high sulfur coal. Results of the first test period will be discussed. Period 1 was used to characterize the operation of the system and determine the preferred process settings for the next test. Period 2 will be a long term test of the dependability of the system while maintaining 90% SO_2 removal and following changes in boiler operation. Plans for Period 3, which will determine how well the system responds to stresses, will also be discussed. Results in the areas of SO_2 collection, limestone utilization, soda ash consumption, and filter cake percent solids have been encouraging. This LSDA demonstration plant is jointly funded by FMC, NIPSCO, the Cleveland Electric Illuminating Company and Southern Indiana Gas and Electric Company. Bechtel Group, Incorporated is evaluating the process for EPRI during Periods 2 and 3.

I. EXECUTIVE SUMMARY

The simplicity and dependability of double alkali scrubbing are now combined with the low cost of limestone regeneration. Together they provide the nucleus for a dependable, cost effective flue gas desulfurization system. These elements have been incorporated in FMC's Limestone Double Alkali (LSDA) demonstration currently being conducted at Northern Indiana Public Service Company's (NIPSCO) Rollin M. Schahfer facility.

This program is being sponsored by FMC, NIPSCO, the Cleveland Electric Illuminating Company (CEI), and Southern Indiana Gas and Electric Company (SIGECO). FMC, with the guidance from all the sponsors, was responsible for the Period 1 testing covered by this paper. The Electric Power Research Institute (EPRI), through Bechtel Group, Incorporated will perform an independent evaluation of the process in subsequent testing.

Period 1 met or exceeded all the performance goals that FMC established for the development program. Throughout the period the system consistently demonstrated SO_2 collection efficiencies in excess of 90% on a midwestern, high-sulfur fuel. In addition, the system produced a dry, easily handled material of 70% to 80% solids which is at least as good as gypsum in its appearance. Limestone utilizations exceeded 90% and sodium losses in the filter cake were shown to be 1.5% or less for two equivalent washes.

II. BACKGROUND

Since 1972, FMC has been installing double alkali FGD systems. These systems use highly soluble salts to react with sulfur dioxide as it is absorbed. Calcium, which is added later externally in a separate reaction tank, exchanges with the sodium to form a solid material out of the sulfur dioxide. The liberated sodium is then reused to collect more SO_2. Lime was chosen as the calcium source in the original systems because it is much more reactive than limestone leading to operating and capital cost benefits. The effectiveness of this system has been proven in a number of installations large and small.[1, 2, 3]

Nevertheless, there has always been a desire to find a way to use limestone to supply the calcium because of its much lower cost and wider availability. The key to its use was to find conditions for the reaction that would allow it to occur at reasonable rates yet not change the scrubbing conditions in any major way. FMC found a way to do just that. Those conditions have been described in other papers[4, 5] so won't be dwelled on here.

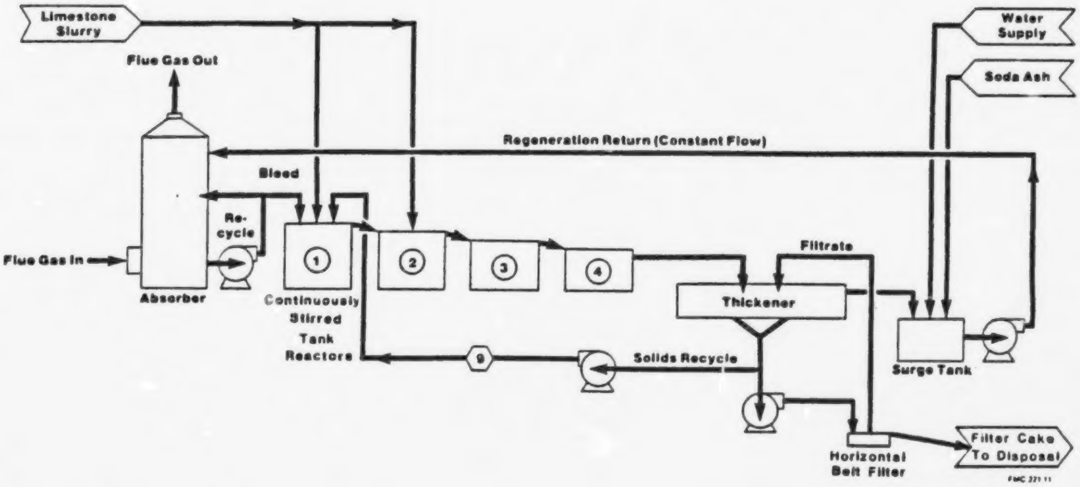
As the first step in demonstrating the concept of LSDA, laboratory pilot tests were run. Those results confirmed our expectations for the process. However, that testing was just for the limestone regeneration system based on a feed of synthesized scrubbing solution. No absorption of SO_2 was attempted, since this was well understood from the lime double alkali experience. The real proof of the process was to come next through a field test of a fully integrated plant.

With the planning and financial help of EPRI, NIPSCO, CEI, and SIGECO, the program for the field demonstration was established. NIPSCO agreed to be the host for the demonstration plant. This was particularly advantageous. Not only do they have flue gas from high sulfur coal combustion and a ready supply of sodium from their 400MW lime double alkali system, but the site is also conveniently located.

The size of the demonstration plant was chosen to be 3MWe. Inlet gas flow is nominally 9,000 acfm. The sulfur dioxide concentration is around 2,500 ppm. About 150 pounds per hour of SO_2 are collected at full load. The size of the plant was chosen to be big enough to use commercial-size equipment which could be simply scaled up for a full size plant, yet small enough to be economical to build and operate. The specific equipment sizes have been described in a prior paper[5], but as an example, the absorber is four feet in diameter and the thickener fourteen feet in diameter. A simplified flow sheet can be found in Exhibit 1.

EXHIBIT 1

LSDA PROCESS FLOW


Limestone Slurry
Flue Gas Out
Water Supply
Soda Ash
Regeneration Return (Constant Flow)
Bleed
Re-cycle
Flue Gas In
Absorber
1
2
3
4
Continuously Stirred Tank Reactors
Filtrate
Thickener
Surge Tank
9
Solids Recycle
Horizontal Belt Filter
Filter Cake To Disposal
FMC 221 11

Testing of the process was set up to be done in three distinct phases following the erection of the plant.

Period 1 - Shakedown Test
Period 2 - Availability Test
Period 3 - Stress Test

The Shakedown Test was designed to characterize the performance of the process over a wide range of operating conditions. From the results of that test, the standard operating conditions for the following tests would be chosen. Period 2 is referred to as the Availability Test or Baseline Test. The intent is to observe how well the system operates over a sixty day run with the only changes being those arising naturally from the boiler operation. The Stress Test planned for Period 3, involves deliberately establishing abnormal input or control conditions to see how the system responds. An example is the improper addition of limestone as would be caused by a malfunction of the pH controller.

Period 1 was completed near the end of February, 1985. Periods 2 and 3 will be run by the end of this year. The purpose of this paper is to present the results of the Shakedown Test.

III. PERIOD 1 RESULTS

The primary goal of Period 1 was to reconfirm the previously described[4] relationships between independent and dependent variables. The factors which affect SO_2 absorption capability, limestone utilization, solids quality, and the ability to control the system would be re-examined under the dynamic conditions of following NIPSCO's boiler. In addition several equipment changes to the basic process scheme needed to be evaluated on a larger scale than previously tested. Among these were a packed absorber utilizing a high efficiency structured packing and an alternate to the stirred tank reactors using the thickener tank as the primary reaction vessel. Equipment variations will be discussed in Section IV.

The independent and dependent variables as well as their relationship to the system design have been covered in previous papers[4, 5] so they will not be re-examined here in depth. Rather we will discuss how the results of Period 1 show the system meets the design objectives FMC set for the program initially. Here then are the design objectives for the program compared to the actual performance during Period 1.

MAINTAIN HIGH SO_2 COLLECTION EFFICIENCY

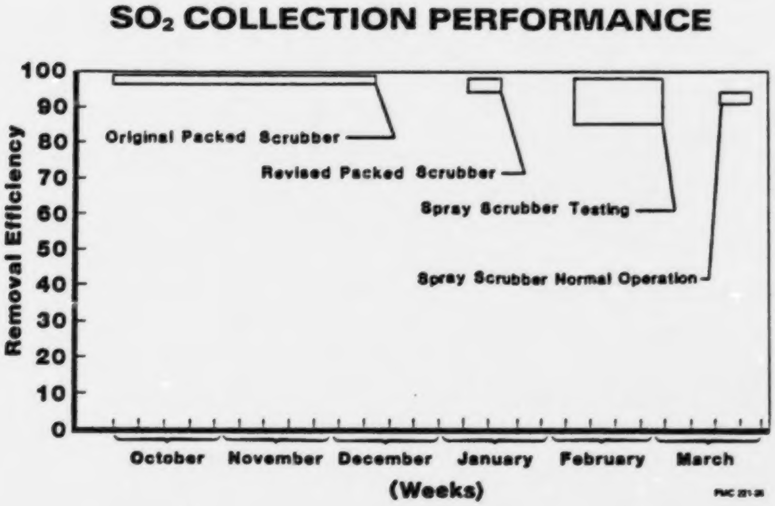
The Period 1 work reconfirmed that the limestone double alkali system can achieve any desired SO_2 collection on high sulfur applications. Collection efficiencies as high as 98% were achieved in both the spray tower and packed tower configurations. This shows that the SO_2 collection capability is dependent primarily on the effective mass transfer area provided.

Exhibit 2 shows the results of SO_2 collection efficiency from October 1984 through March 1985. The work on SO_2 collection efficiency can be divided into roughly 4 periods.

The first period was with the original packed tower design. This routinely gave collection efficiencies in the high 90's as can be seen. Unfortunately, the gas side pressure drop was higher than anticipated.

For a variety of reasons, which are discussed in the section dealing with the packed tower, the flow schematic for the absorber was changed from one in which the regenerated liquor was pumped directly to the top of the packing to one in which the regenerated liquor was blended in the absorber sump prior to being pumped up to the packing. This meant that the sump liquor reduced the pH and the collection efficiency was consequently reduced slightly as can be seen in Exhibit 2 for January of 1985.

EXHIBIT 2



Still not completely satisfied with the performance of the packed tower, we decided to convert the demonstration plant absorber to a spray tower. Since the geometry of the tower was fixed, combinations of distribution plate open area, L/G, pH, etc. were tested during February in an attempt to bracket the performance. Again SO_2 collection efficiencies in the high 90's were achievable. After discussions with the utility sponsors it was decided to fix the absorber design to do 91 to 94% collection efficiency. This represented a compromise between those who wished to demonstrate as high a collection as possible and those who felt that exactly meeting the NSPS requirement of 90% was all that was required.

The final configuration selected for the spray tower can be seen in Figure 1. Wall affects due to the small absorber diameter contribute to the relatively higher L/G for this system over what a larger scale spray tower would need.

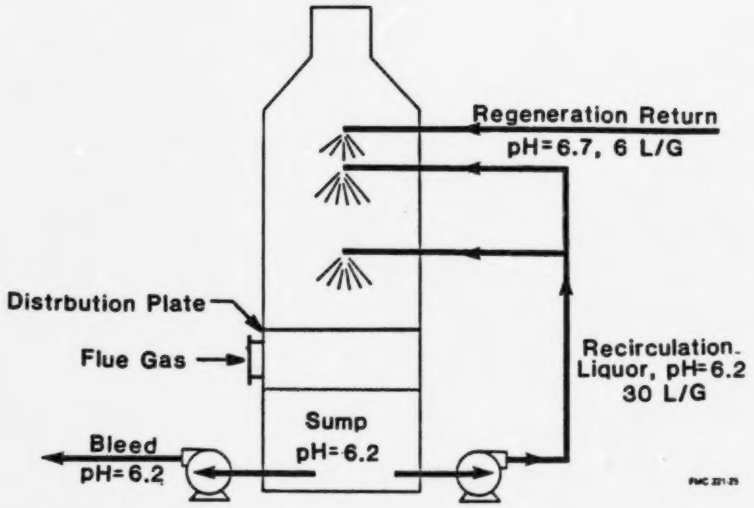

Regeneration Return
pH=6.7, 6 L/G
Distrbution Plate
Flue Gas
Recirculation.
Liquor, pH=6.2
30 L/G
Sump
pH=6.2
Bleed
pH=6.2

Figure 1. Spray Scrubber Liquor Flow Scheme

The results since March have routinely demonstrated SO_2 collection efficiencies of 91 to 94% collection at only 2.0 to 3.0 inches total gas side pressure drop and an L/G ratio of 35.

One of double alkalis advantages in maintaining SO_2 collection efficiency was demonstrated by preliminary stress tests performed during Period 1. To show the stability of SO_2 removal, as well as the systems insensitivity to reagent feed, each reagent was shut off and the SO_2 removal was monitored.

The limestone feed was turned off completely and it was a full 12 hours at full load before the system dropped below 90% removal efficiency. It was an additional 9 hours before the removal dropped below 81% collection efficiency.

The systems soda ash supply was shut off for 20 hours with no measureable difference in SO_2 collection. The soda ash test was stopped at this point since calculations showed it would be literally weeks before SO_2 collection would be affected.

In attempting to understand this phenomenon one needs to understand that the sodium based double alkali system (either limestone or lime) does not rely on either reagent directly to collect SO_2. Any single alkali system, such as lime/limestone or even sodium scrubbing is dependent on the small amount of available alkali in the scrubber system proper. The double alkali system collects SO_2 with the soluble sulfite inventoried in the regeneration system. The calcium merely converts bisulfite over to the higher pH sulfite. The soda ash merely makes up for system losses. Since the inventory of soluble sulfite is quite large changes in the reagent supply are slow to change this inventory.

One additional point is noteworthy. While other testing pertaining to limestone utilization, solids quality, tramp materials and alternate limestones were being performed the system was capable of continually exceeding the minimum 90% collection efficiency.

Achieve A Minimum of 90% Limestone Utilization

During Period 1 the system demonstrated excellent limestone utilizations. The system can routinely meet or exceed 90% limestone utilization. A 90% utilzation of limestone is equivalent to a limestone stoichiometric ratio of 1.07 moles $CaCO_3$ per mole of SO_2 collected for a double alkali system that has 3.5 mole % soda ash consumption. A 95% utilization would be a 1.02 stoichiometric ratio.

Two of the more important variables from the standpoint of limestone utilization are limestone grind and reactor residence time. The reactor residence time for the demo plant was fixed during the design of the plant. The demonstration plant reactor residence time is 180 minutes. While this number appears large at first glance it is very comparable to the tankage required for a conventional limestone system. This reactor residence time is expressed on a different basis than those used by the conventional limestone suppliers which causes some confusion. The LSDA residence time is based on the bleed rate from the absorbers while with limestone systems the reference to recirculation tank residence time is in terms of absorber L/G rather than bleed. For comparison, lets assume 94% collection on a 3.5% sulfur coal. The LSDA system would have 180 minutes residence time based on a bleed flow rate equivalent to 6 L/G giving 6 x 180 minutes or 1080. For the limestone system this would be equal to a 12 minute tank based on a L/G of 90. The advantage of double alkali is that this tankage can be located remotely from the flue gas absorbers which can be extremely important for sites with limited space as might be encountered in retrofit situations.

The limestone grinding system was not provided for in the demo plant to save on capital. The plant therefore receives pre-ground limestone into a dry storage bin by pneumatic truck. From there it is slurried and pumped to use points in the plant. The only disadvantage to this is that one is limited to commercially available dry ground material. Two of the limestones used for Period 1 are characterized in Table 1. While it would be desireable to obtain material which is 95% - 325 mesh for comparison purposes to other systems the commercially available pre-ground stone appears to be either 92% - 325 or 99% - 325 mesh.

TABLE 1. LIMESTONE ANALYSIS AND GRIND FOR PERIOD 1

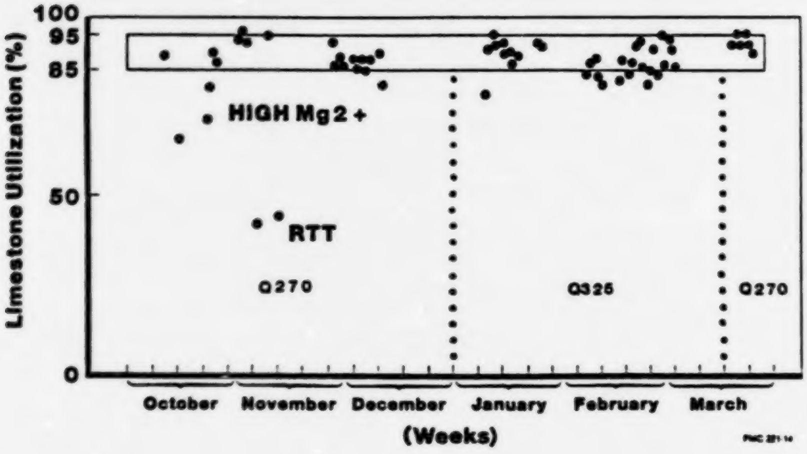
	Hubercarb™ Q270	Hubercarb™ Q325
$CaCO_3$ %	96.0	96.0
$MgCO_3$ %	2.2	2.2
Inerts %	1.8	1.8
Size Distribution:		
% 200 mesh (74 microns)	100	100
% 230 mesh (63 microns)	99.5	100
% 270 mesh (53 microns)	97	99.9
% 325 mesh (44 microns)	91	99

Note: 1. Actual chemical analyses of limestone is reported.
2. Size distribution is based on actual Coulter Counter data.

Exhibit 3 summarizes the chronology for limestone utilization over the first 6 months of the program. These utilizations are based on residual carbonate in the filter cake. This plot has not been corrected for residual magnesium carbonate. If corrected, values would be approximately 2 points higher.

EXHIBIT 3

LIMESTONE UTILIZATION



For the first 3 months of the program the Hubercarb^{TM} Q-270 material was used. In general, the only poor utilizations over this time period were recorded when the system was artificially spiked with soluble magnesium and during operation of the RTT.

Magnesium interferes with the limestone utilization. The natural magnesium concentration from the use of Hubercarb^{TM} stone was 300 ppm. The level was raised to 800 ppm by the addition of magnesium chloride. The limestone utilization dropped from approximately 90% to around 70%. The filter cake remained stable at approximately 75% solids. These high solids may have been due in part to the unreacted limestone particles in the filter cake. To counteract the reduction in utilization, operating conditions were modified and the pH reduced. As in our previous bench scale work the utilization climbed back up to its former level but the filter cake percent solids went down. This points out another aspect of the systems flexibility since the designer can optimize either solids quality or limestone utilization depending on which conditions are economically favorable. Once the magnesium addition to the system was stopped, the concentration dropped back to its historical level as the magnesium was worked out of the system. The system operation returned to normal. During this entire period the systems SO_2 collection remained in the high 90's.

From January through early March the Q325 material was used in the pilot plant. Overall, the Q325 material raised the utilizations 3 to 5 percentage points. Low utilization in February was a result of less favorable chemistry conditions while spray scrubber operation was bracketed. Additionally, during this period, it was discovered that at least one load of limestone had some of the smaller size fraction material removed prior to shipment. Upon learning of this we soon rectified the situation with the supplier.

At the end of March, the limestone supply was returned to the Q270 material.

Maintain/Improve The Quality Of The Waste Products

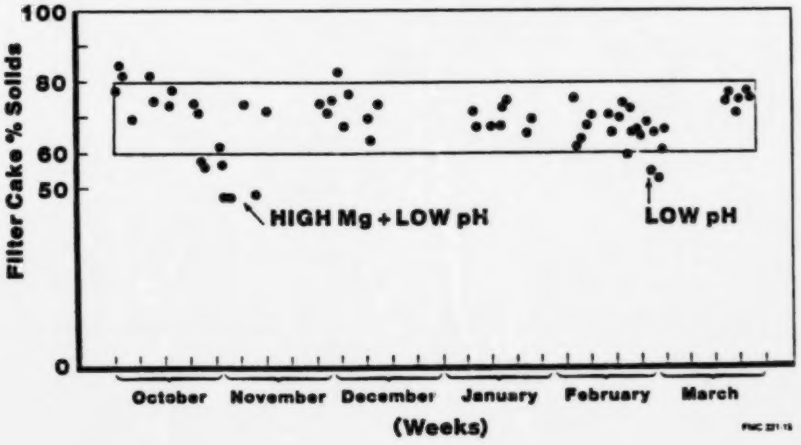
The quality of the filter cake from the system has been outstanding. The filter cake percent solids for the limestone double alkali demo directly parallel those of the Unit 17 lime double alkali system. Both systems produce 70 to 80% solids materials which are dry and crumbly. Exhibit 4 shows the chronology for filter cake solids during Period 1. The only notable exceptions to the excellent filter cakes was found during the high magnesium level testing when the pH was reduced to increase utilization previously mentioned. The other low filter cake solids period occurred during a low pH period designed to bracket the spray tower performance.

It is of special interest to note that during the entire Period 1 testing the problems of non-settling solids experienced by EPA during its limestone dual alkali testing at Plant Scholz[7] did not occur. The only occurence of thickener overflow clarity problems occured during RTT testing and will be covered under that section.

Another benefit to the higher solids filter cake is the improved sodium recovery. The double alkali filter cake is washed to recover the valuable sodium salts from the filter cake so that they may be re-used in the absorber. The demo plant uses a horizontal filter with counter current

EXHIBIT 4

FILTER CAKE QUALITY



wash. Washing efficiency is normally expressed in terms of an equivalent wash which is equal to the amount of solution remaining in the filter cake after filtration. The sodium consumption as analyzed in the filter cake was 1.5 moles of Na_2CO_3 per 100 moles of SO_2 collected for two equivalent washes. For one equivalent wash the consumption was 2.8 mole percent. Even with no wash water the consumption was only 5.1 mole %.

Maintain The Control Advantages Of Lime Double Alkali

The pilot plant is controlled by using FMC's patented control system (U.S. Patent Nos. 4,431,618). This control scheme estimates the amount of SO_2 collected by measuring the process variables (flow, pH, and active sodium concentration). This value is used to pre-estimate the amount of limestone required. The control algorithm has been programmed onto an IBM^{TM} personal computer which estimates the limestone feed rate for the operator. The system does not rely on maintenance intensive SO_2 monitors and reacts automatically (that is, without SO_2 monitors) to changes in fuel sulfur content, firing rate, and/or absorber collection efficiency.

For a great percentage of the time during Period 1, the demonstration plant was operated in a manner which simulated a swing load. The absorber was operated for full load 12 hours per day and reduced to half load at night.

IV. EQUIPMENT MODIFICATIONS

Two major equipment modifications were tested during Period 1. These were an absorber tower filled with a structured packing and a thickener alternate which functioned as the primary reactor for calcium sulfite precipitation. The packed tower showed sufficient promise to warrant futher testing later in or subsequent to Period 3. The reactor thickener tank (RTT) showed little promise.

PACKED TOWER

From a mass transfer standpoint the double alkali system is extremely simple. This system is only dependent on gas phase resistance unlike direct calcium systems which may be gas phase and liquid phase dependent. This means that for any given case the SO_2 collected is a function of the mass transfer area available. This is represented by the amount of liquid surface available to absorb SO_2. Once the design criteria (such as flue gas composition, SO_2 collection required, etc.) have been established the required mass transfer can be generated in many different ways.

Some absorber designs shear the absorber liquid into many droplets by accelerating the flue gas through the liquid (i.e. trays, disc and donuts, etc.). This method relys on fan or gas side horsepower to generate sufficient surface area.

The preferred method today in the utility industry is to accelerate the scrubbing liquid through nozzles which generate a large number of tiny droplets. This is the spray tower design which relies on liquid side horsepower (nozzle ΔP and higher L/G ratios) to generate the required mass transfer surface. Hybrid designs which utilize both methods are also common.

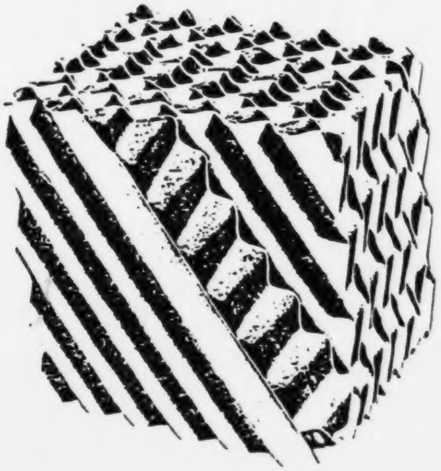
A third option is used quite successfully in the chemical industry although it is not common in the utility FGD market. This is the use of structured packings. In this device liquid flows down a structure which has a very large surface area to volume ratio. If this ratio is high enough a large amount of intimate gas/liquid contact occurs as the flue gas flows past these thin sheets of wetted packing. Since the mass transfer area is generated by the structured packing itself this option represents a minimum for L/G and gas side pressure drop.

Choosing a specific device for absorption involves evaluating trade-offs between fan energy, pumping energy, equipment capital cost, and dependability/maintainability features. Based on previous work with lime double alklai, 90% or greater SO_2 collection could be achieved with the packing on high sulfur coal with a liquid requirement of less than 6 gal per thousand cubic feet of gas, a lower gas side pressure drop than an equivalent spray tower and a significantly lower tower height. This provides substantial capital and operating advantages.

The only unknown was its long term dependability of operation under normal plant conditions. The passageways in the most efficient packing are relatively narrow (as small as 1" x 1/2" in a diamond shape) and could be susceptible to blockage should large amounts of fly ash or other solids get into the absorber. The only way to determine if this was a problem was to operate with it. Thus we chose to begin Period 1 testing with the packed absorber. The structured packing consisted of many sheets of corrugated material fastened so that the corrugations are oriented at an angle to one another. (See Exhibit 5)

EXHIBIT 5

STRUCTURED PACKING

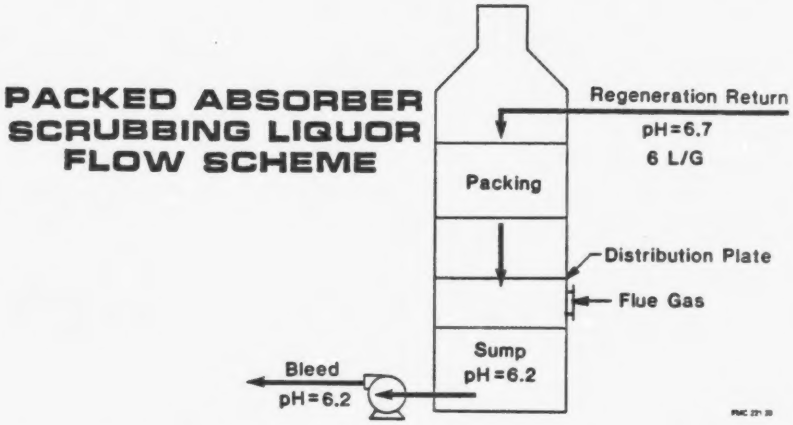


The sulfur dioxide collection performance of the packed absorber was excellent. Removal was 95% to 99% over a wide range of conditions. Disappointingly, the pressure drop was higher than expected initially and increased during operation due to the deposition of solids on the surface of the packing. Because of this and the need to move ahead with the scheduled program, we switched to a spray absorber for the Period 2 and 3 tests. After examining all the evidence, however, we feel these drawbacks can be corrected or minimized and expect to test the packed absorber again after Period 3. This evidence is discussed below.

From our previous testing, which was conducted without a distribution plate, we expected the pressure across the packing to be about 2 inches w.g. It was closer to 3. Measurements indicated the excess was across the packing rather than the distribution plate.

The distribution plate, which was located below the packing resulted in very high velocities through the plate. (See Exhibit 6) Even at the low

EXHIBIT 6


PACKED ABSORBER SCRUBBING LIQUOR FLOW SCHEME
Regeneration Return
pH=6.7
6 L/G
Packing
Distribution Plate
Flue Gas
Sump
pH=6.2
Bleed
pH=6.2

liquid to gas rates being run, a substantial amount of SO_2 collection was occuring on the distribution plate. Due to the small size of the demo plant absorber, there was insufficient height to effectively de-entrain the liquor from the gas prior to entering the lower portion of the packing. A greater open area on the plate will aide in reducing the high liquid loading to the bottom packing. Also, larger size absorbers will have additional disengaging height above any distribution baffle.

We first noticed pressure drop increases in the absorber during the RTT operation. At that time, it was assumed that the higher carryover of suspended solids from the RTT was the cause. After cleaning the packing and returning to the Continuosly Stirred Tank Reactor (CSTR) operation, we again experienced increasing pressure drops. We found that solids had deposited on the packing surface in a thin layer. In some places, this layer would flake off, and the flakes seemed to get caught in the passageways of the packing.

In our prior tests of packing we had not seen evidence of such deposition. One difference was the direct feed of regenerated scrubbing solution to the packing before this stream was mixed with solution in the absorber sump which was then recirculated to the packing. Conversion to this type of operation brought improvements, but there was still some increase in pressure drop.

When the packing was removed from the absorber for conversion to the spray tower, it was noted that the pluggage occurred mainly in those passageways which terminated at the wall of the absorber. The flakes of material seemed to pass through the packing until they hit a wall then built up there. The center of the packing had the thin layer of deposits but essentially no plugging. In a small diameter vessel like this, the number of passageways ending at a wall comprise a disproprotionately large percentage of the area compared to a commercial size absorber. This means that the pressure drop of a larger size absorber will not be affected nearly as much.

In future testing of the packed absorber, we will try packing with more open passageways, modify the distribution plate, and try to eliminate the absorber wall/packing interface problem. We still have high expectations that the packed absorber will prove to be the preferred scrubber alternative.

REACTOR THICKENER TANK (RTT)

The RTT was investigated as an alternate to the continuously stirred tank reactors. It was hoped that the RTT would represent considerable capital savings in addition to providing additional crystal surface area to improve solids quality.

In theory, the RTT obtains the necessary retention time by having the solids remain in the reaction zone for a longer time period than the liquid. This results in the same residence time for a smaller vessel. This separation (and differential residence time) results in an increased solids concentration in the reaction zone. Thus the reaction and settling occur in the same vessel. Similar equipment is used in other applications within the process industry.

The center portion of the RTT is a turbulent reaction zone, which is separated by a cylindrical baffle from the outer, quiescent settling zone. The bleed solution and limestone slurry are fed into the reaction zone. An agitator and draft tube provide internal recirculation of the limestone and calcium sulfite particles, maintaining a slurry concentration of 3% or greater. Eventually the calcium sulfite, and some unreacted limestone falls to the bottom of the tank, where it thickens to a 20 percent slurry and is pumped to the filter. Clear supernatant overflows the annular, quiescent zone of the RTT.

Unfortunately, the RTT did not perform well. The solids could not be kept in the reaction zone for the desired residence time. This led to poor utilization of the limestone and poor settling of the solids. Utilization went down as low as 50%. Filter cake quality remained good but that was most likely because of the high amount of unreacted limestone present.

Of even greater concern was that the solids were not settling as they did in the CSTR operation. Instead of 100 ppm suspended soids in the overflow liquor we were getting up to 1,000 ppm. Since the overflow liquor is pumped back to the absorber, all of those solids were being sent there too.

At the time, we were using the packed absorber and the solids were being deposited on the packing. This caused the pressure drop across the absorber to increase a couple of inches water gauge after only a few days operation of the RTT.

During our testing there was no discernable improvement in filter cake properties when using the RTT versus the four stirred tanks. In addition, the difficulty one would encounter in monitoring and controlling the RTT made it impractical for power plant use.

Recent cost estimates have shown that there is little, if any, cost advantage for the RTT approach. The agitation and the baffles necessary to utilize a single reaction vessel approach the cost of the extremely simple stirred tanks.

For these reasons the RTT operation was terminated. Even before testing began the high quality of solids produced by the stirred tanks reduced the incentive to use the RTT. The poor performance of the RTT during testing made it clear that the stirred tank approach should be used for the remainder of the formal program.

V. FUTURE TESTING

Period 2 of the test program is a long term test of the performance and dependability of the systems. Settings of the operation will be selected based on the Period 1 results. The system will be run for 60 days at those conditions with the only changes being those which arise normally out of the boiler operation. EPRI, through Bechtel Group, Inc., will monitor and evaluate the performance, including reagent and power consumption, waste generation and quality, collection efficiency, and dependability.

Period 3 will consist of a series of stress tests. That is operating conditions will deliberately be altered to see how the system responds to the change and to corrective action. Such tests are expected to provide an indication of the insensitivity of the system to operator error, equipment malfunction, changes in reagent supply, or changes in boiler operation. The test plan for this period is not complete but included will be tests of limestone compositions, water quality, and pH control. Bechtel will be evaluating these tests also.

All testing is to be complete by the end of 1985. we expect that the results from Periods 2 and 3 will provide independent confirmation of those experienced in Period 1 and reported here. The Period 1 results show that LSDA is unsurpassed in its ability to dependably and economically collect sulfur dioxide at high efficiency from high sulfur coal.

The work described in this paper was not funded by the U. S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no offical endorsement should be inferred.

BIBLIOGRAPHY

1. Coons, John D., Kevin C. Rorke, and Clifford A. Rosene, "Double Alkali Operating Experience at the NIPSCO R. M. Schahfer Generating Station," presented at ASME's 1984 Industrial Power Conference, Toronto, Ontario, September 30 to October 4, 1984.

2. Van Meter, James A., Thomas H. Durkin and L. Karl Legatski, "Operating Experience with the FMC Double Alkali Process," presented at the EPA Symposium on Flue Gas Desulfurization, Houston, Texas, October 29, 1980.

3. Francis, D. V., R. J. Biolchini, and J. D. Coons, "Operating Experience with High Sulfur Coal in an Industrial Double Alkali FGD System," presented at the Coal Technology '81 Conference, Houston, Texas, November 17-19, 1981.

4. Boward, W. L., R. O. Petkus, and K. H. Wang, "FMC Limestone Double Alkali FGD Process," presented at the American Institute of Chemical Engineers, Summer National Meeting, Cleveland, Ohio, August 30, 1982.

5. Biolchini, Robert J. and Beth A. Wrobel, "EPRI/NIPSCO/FMC Limestone Double Alkali Demonstration Plant," presented at the Air Pollution Control Association 77th Annual Meeting, San Francisco, California, June 24-29, 1984.

6. Petkus, R. O., Robert J. Biolchini, "Current Progress at FMC Limestone Double Alkali Demonstration Plant," presented at the Southeastern Electric Exchange, Atlanta, Georgia, April 11, 1985.

7. Valencia, J. A., and Peirson, J. F, Jr., "Evaluation of the Limestone Dual Alkali Prototype System at Plant Scholz: Final Report," U.S. EPA Report, EPA-600/7-81-146, 1981.

EFFECTS OF LIMESTONE TYPE AND GRIND
ON
DUAL ALKALI SYSTEM PERFORMANCE

By

John C. S. Chang
Acurex Corporation
Southeast Regional Office
P. O. Box 13109
Research Triangle Park, NC 27709

Theodore G. Brna and Norman Kaplan
U. S. Environmental Protection Agency
Air and Energy Engineering Research Laboratory
Research Triangle Park, NC 27711

ABSTRACT

Pilot plant tests were conducted to evaluate the dual alkali system performance with respect to five types and grinds of limestones. The data show that high purity, high calcium, fine grind limestone should be used for the regeneration of dual alkali scrubbing solution. Limestone containing more than 0.5% magnesium is not suitable for dual alkali processes. The accumulation of dissolved magnesium caused performance deterioration and system upset. Experimental techniques which can be used to characterize the limestone reactivities were also evaluated. The reactivities measured by sodium bisulfite titration agreed with pilot plant data.

INTRODUCTION

Limestone is a preferred chemical reagent for dual alkali (DA) flue gas desulfurization (FGD) processes due to its abundance and low cost. The most commonly used limestone is naturally occurring carbonate rock or fossil matter which is composed primarily of calcium carbonate or a combination of calcium and magnesium carbonates. There are a great many different forms and types of limestones, varying in color, chemical composition, mineralogy, crystallinity, texture, and hardness.(1) When selecting limestone for dual alkali FGD processes, variables which should be considered include physical characteristics, chemical requirements (impurity tolerances), and size or gradation of the pulverized (or ground) stone.

Physical characteristics can affect limestone grindability and grinding circuit energy requirements. The relationship between limestone grindability and the economics of FGD reagent preparation has been discussed by Colley et al.(2) Correlations were developed for estimation of ball mill size, specific energy, and production rate. On the other hand, little work has been done on chemical compositions and grinds (particle size distributions) of limestones, both of which have the potential of affecting dual alkali process chemistry, system performance, and operability. The significance of these potential effects can be illustrated by the results of the limestone type and grind tests on other (slurry type) FGD processes(3,4,5). Significant increases in SO_2 removal efficiency and reagent utilization were achieved by using fine grind (90% less than 325 mesh) limestone. Decreases of scaling potential were reported by choosing the appropriate type and grind of limestone. Increased SO_2 absorption also resulted with the soluble magnesium in the limestone.

In order to support future full-scale FGD operations and broaden the dual alkali data base, the U.S. Environmental Protection Agency's (EPA's) Air and Energy Engineering Research Laboratory (AEERL) at Research Triangle Park, NC, sponsored a series of pilot plant tests with the objectives of:

- evaluating the effects of limestone type and grind on sodium based dual alkali system performance, and
- developing guidelines for selection and specification of limestone for dual alkali processes.

This paper summarizes the test results and discusses laboratory procedures which can be used to aid limestone selections.

TYPES AND GRINDS OF LIMESTONES TESTED

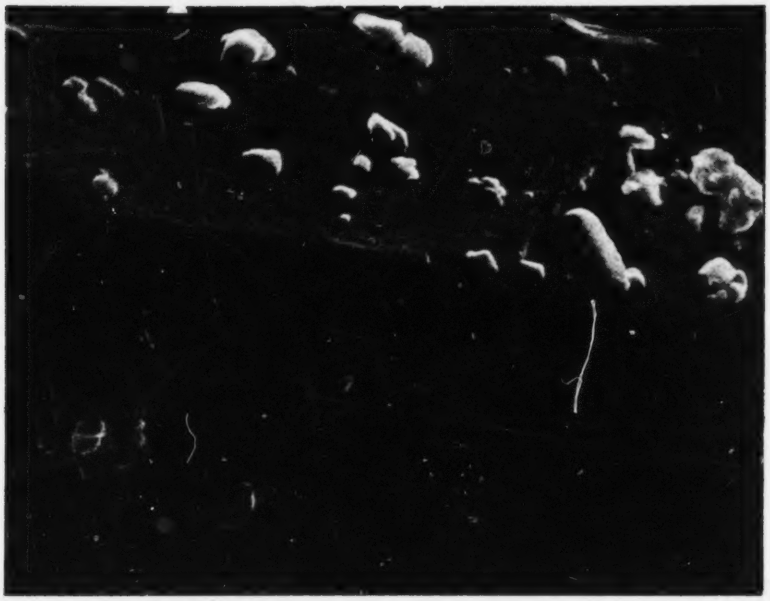
Five types of limestones were tested in the RTP dual alkali scrubber. Typical chemical composition of each type of limestone is shown in Table 1. Three high calcium limestones (Aragonite, Sylacauga, and Fredonia), each containing less than 0.4% magnesium, were tested to compare the calcite (rhombohedral crystal) with the Aragonite (orthorhombic crystal). To evaluate the effect of magnesium in limestone on dual alkali system performance, Tate (0.8 to 1.0% magnesium) and Whitestone (1.6 to 4.0% magnesium) limestones were tested.

The surface structures of the ground limestones were examined by a scanning electron microscope (SEM). Limestone particles with surfaces ranging from flat and smooth to porous and fibrous were obtained with Sylacauga and Aragonite as two extremes. The crushed Sylacauga limestone consists of fragments of milky white to gray rhombs as shown in Figure 1(a). The particles have distinctly curvilinear boundaries and smooth surfaces. The shapes of the particles are highly irregular with sharp edges. On the other hand, Aragonite limestone is more amorphous as shown in Figure 1(b). The exterior surfaces of the particles are fibrous as revealed by the high magnification (x10,000) SEM picture in Figure 1b. Fredonia limestone has a surface structure intermediate to that of

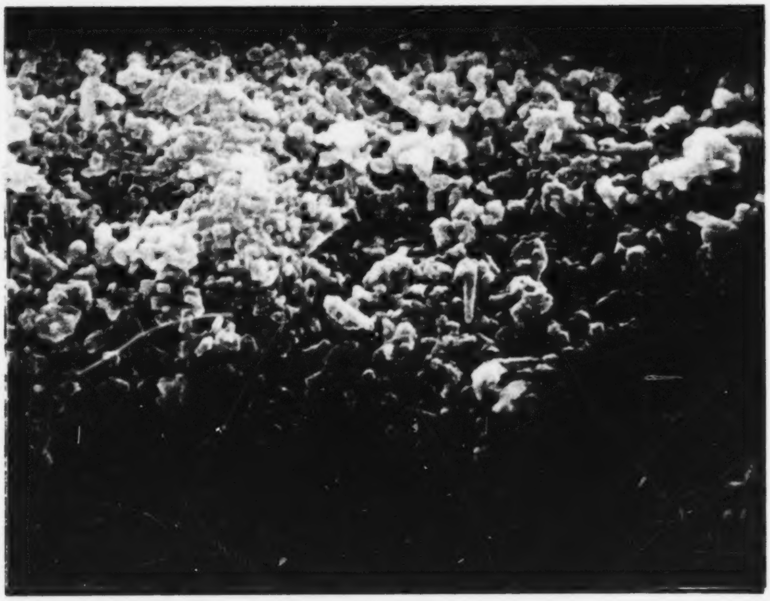
TABLE 1. TYPICAL COMPOSITIONS OF LIMESTONES TESTED

Limestone	Source	Location	Chemical Composition (mg/g) Ca	Mg	CO_3	Fe_2O_3	SiO_2	Al_2O_3
Aragonite	Marcona Ocean Industries, Ltd.	Bahama	390 (min)	2-3	595 (min)	ND[a]	1.5-3.5	2.5-3
Sylacauga	Georgia Marble Company	Sylacauga, Alabama	390 (min)	2.5-3.5	595 (min)	0.2-1.6	2-14	1-3.5
Fredonia	Fredonia Quarries	Fredonia, Kentucky	390 (min)	3-4	595 (min)	0.5-1.4	2.5-3.5	4.5-5
Tate	Georgia Marble Company	Tate, Georgia	370 (min)	8-10	570 (min)	0.2-1.6	2-14	1-3.5
Whitestone	Georgia Marble Company	Whitestone, Georgia	340 (min)	16-40	570 (min)	0.2-1.6	2-14	1-3.5

[a] ND - not detected



(a) SYLACAUGA



(b) ARAGONITE

Figure 1. Scanning electron micrographs of (a) Sylacauga and (b) Aragonite at 10,000 x magnification.

Sylacauga and Aragonite in terms of particle shape and surface smoothness. Tate and Whitestone are similar to Sylacauga.

Limestone grinds are characterized by particle size distributions measured by wet screen analysis. Five nominal grinds (99% less than 325 mesh, 90% less than 325 mesh, 90% less than 200 mesh, 90% less than 140 mesh, and 90% less than 80 mesh) were used to represent the extent of limestone grinding.

In addition to size distribution, another physical property of limestone particles measured is the surface area. Surface areas were determined using the multipoint Brunauer, Emmett, and Teller (BET) method. Results are listed in Table 2. The large surface area of Aragonite reflects the porous surface structure shown by SEM pictures. Fredonia has less surface area than Aragonite but still considerably more than Sylacauga, Tate, and Whitestone. However, for each type of limestone, no significant differences in BET surface areas among various grinds were detected.

APPROACH

The scrubber configuration shown in Figure 2 was used for the limestone type and grind tests. The SO_2 scrubber is a three-stage tray tower with 7.5 m^3/min. (approximately 0.1 MW) flue gas capacity. Flue gas is drawn from a gas-fired boiler (no flyash is present); pure SO_2 is injected to achieve the desired concentration in the flue gas. Regeneration of spent scrubbing liquor is performed in the four-tank-in-series reactor train with a total residence time of 80 minutes. Limestone is fed to the first reactor as 45% slurry, with the feed rate manually set as required for either pH or reagent stoichiometry control. Soda ash is added to the fourth reactor as sodium makeup. The reactor effluent slurry flows by gravity to the thickener centerwell. Clarified liquor overflows from the thickener to the forward feed hold tank from which it is pumped to the tray tower. A horizontal belt filter is used for further dewatering of the thickener underflow solids.

TABLE 2. LIMESTONE PARTICLE SURFACE AREAS MEASURED BY BET[a] METHOD

Type	Grind[b]	Surface area (m^2/g)
Aragonite	90%-325	4.61
Aragonite	90%-200	4.13
Aragonite	90%-140	4.45
Aragonite	90%-80	3.58
Sylacauga	99%-325	0.73
Sylacauga	90%-325	0.66
Sylacauga	90%-200	0.42
Sylacauga	90%-140	0.47
Fredonia	99%-325	2.37
Fredonia	90%-325	2.21
Fredonia	90%-200	1.95
Whitestone	90%-325	0.88
Tate	99%-325	0.72
Tate	90%-325	0.77
Tate	90%-200	0.63
Tate	90%-140	0.47

(a)Brunauer, Emmett, and Teller.
(b)Weight percentage passing through screen of indicated mesh.

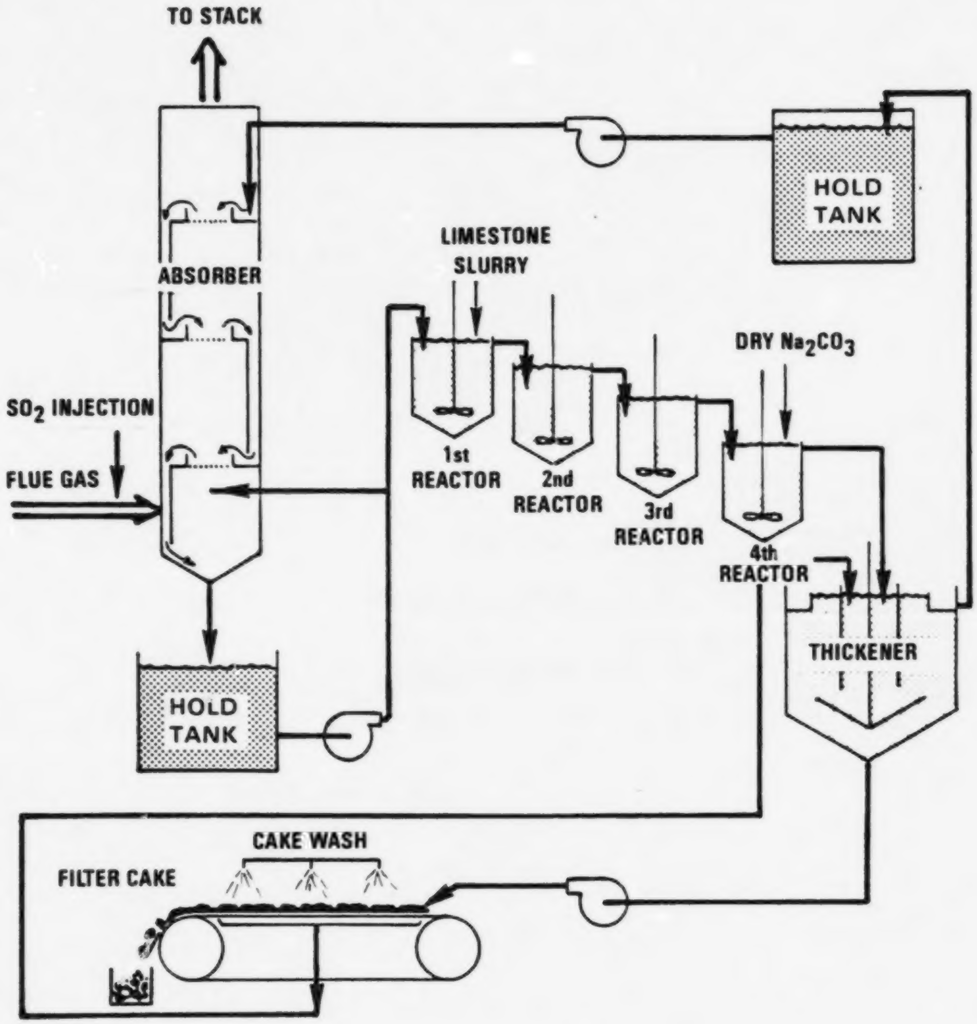

TO STACK
ABSORBER
HOLD TANK
LIMESTONE SLURRY
DRY Na_2CO_3
SO_2 INJECTION
FLUE GAS
1st REACTOR
2nd REACTOR
3rd REACTOR
4th REACTOR
THICKENER
HOLD TANK
CAKE WASH
FILTER CAKE

Figure 2. Flow diagram of RTP limestone dual alkali pilot plant.

Major operating conditions are listed in Table 3. The operating conditions were maintained constant so that limestone type or grind was the only variable for the test series.

Ultraviolet spectrophotometry (DuPont 400 SO_2 analyzer) was used to monitor the gas phase SO_2 concentrations at the inlet and outlet of the scrubber to determine SO_2 removal efficiencies. The pH of the scrubbing liquor in each reactor was measured hourly during the pilot testing. Solids dewatering properties were characterized by the hold tank slurry settling rate and filter cake insoluble solids concentration. Limestone stoichiometry was determined by mass balance and filter cake analysis. Detailed descriptions of the test facilities and analytical procedures have been reported earlier(6).

RESULTS AND DISCUSSION

Pilot plant data for five limestones demonstrated that limestone type and grind have significant effects on system performance parameters which include:

- o System pH
- o SO_2 removal
- o Limestone stoichiometry*
- o Solids settling rate
- o Filter cake insoluble solids concentration

Filter cake quality also affected liquor losses and soda ash consumption.

A wide range of results was observed in the testing. For example, for the most reactive limestone (90% less than 325 mesh Aragonite), filter cake insoluble solids exceeded 60%; the pH of regenerated liquor reached 6.8; SO_2 removal efficiency attained 94% with 1.0 limestone stoichiometry; and solids

---

*Limestone stoichiometry = moles of $CaCO_3$ added/moles of SO_2 absorbed

TABLE 3. OPERATING CONDITIONS OF DUAL ALKALI LIMESTONE TYPE AND GRIND TESTS

Parameter	Value
Liquid-to-gas (L/G) ratio	0.8-1.07 L/m^3 (6-8 gal./1000 acf)
Limestone slurry solid content	45%
Flue gas SO_2 concentration	3000 ppm
Flue gas O_2 concentration	7-8%
Active sodium in scrubbing solution	1.0 M
Number of reactors	4
Total reactor residence time	80 min.

settling rate also reached 2.2 cm/min. On the other hand, for the least reactive limestone tested (90% less than 325 mesh Whitestone), insoluble solids dropped to 30%; the pH of regenerated liquor was only 5.9; SO_2 removal efficiency was as low as 76% with 1.3 limestone stoichiometry; and non-settling solids (settling rate less than 0.1 cm/min) were produced which caused system upset, plugging, and scaling. Detailed results of each limestone tested are discussed below.

Sylacauga

Five runs were made with four different grinds of Sylacauga limestones; the principal results are listed in Table 4. These test results indicated that, when maintaining constant limestone feed rate, the pH of regenerated liquor decreased, causing a slight drop in SO_2 removal efficiency when coarser grinds of Sylacauga were used (runs SY-3 and SY-5). To maintain constant pH and SO_2 removal, the limestone stoichiometry had to be increased. Limestone feed stoichiometry of 1.01 was used to obtain a regenerated liquor pH of 6.7 with the finest ground Sylacauga, 99% less than 325 mesh (run SY-1). No significant change in stoichiometry was required to maintain the regenerated liquor pH at 6.6 with the second grind, 90% less than 325 mesh (run SY-2). But, when the 90% less than 200 mesh Sylacauga was used, a 1.11 stoichiometry was required to attain a pH of 6.6 (run SY-4).

In addition to stoichiometry changes, the most significant result obtained was the effect of limestone grind on filter cake quality. Insoluble solids, 54% by weight, were obtained in the filter cake when using 99% less than 325 mesh Sylacauga (run SY-1). The filter cake insoluble solids dropped to 48% when the coarser Sylacauga (90% less than 325 mesh) was used (run SY-2). Runs SY-3 and SY-4 showed that only 43 to 44% solids in the filter

TABLE 4. RESULTS OF DUAL ALKALI PILOT TESTS WITH SYLACAUGA

Run No.	SY-1	SY-2	SY-3	SY-4	SY-5
Nominal grind, mesh	99%-325	90%-325	90%-200	90%-200	90%-140
Scrubber feed pH	6.7	6.6	6.4	6.6	6.3
Scrubber effluent pH	6.3	6.2	6.1	6.2	6.0
Inlet SO_2, ppm	3030	3050	3020	3050	3040
SO_2 feed rate, kg/hr	2.9	2.9	2.9	2.9	2.9
SO_2 absorption, %	94	92	91	92	90
Limestone slurry feed rate, kg/hr	9.5	9.5	9.5	10.5	9.5
Flue gas O_2, %	7.8	8.1	7.7	7.1	8.1
Filter cake solids, % (insoluble)	54	48	44	43	42
Total alkalinity, gmol/L	0.4	0.38	0.35	0.37	0.33
Run time, hr	78	82	8.1	88	86
Limestone stoichiometry	1.0	1.01	1.03	1.11	1.05
Settling rate (reactor), cm/min.	2.9	2.8	2.5	2.4	2.4
Settling test % solids	2.4	2.3	2.3	2.5	2.3
Active Na, gmol/L	1.03	1.0	1.01	1.00	0.98
SO_4/SO_3	1.94	1.98	1.99	2.0	2.2

Cake was obtained when 90% less than 200 mesh Sylacauga was used. A further decrease in filter cake insoluble solids, to 42%, was obtained with the coarsest grind (run SY-5). A slight drop of solids settling rate was also observed with the coarser Sylacauga.

Aragonite

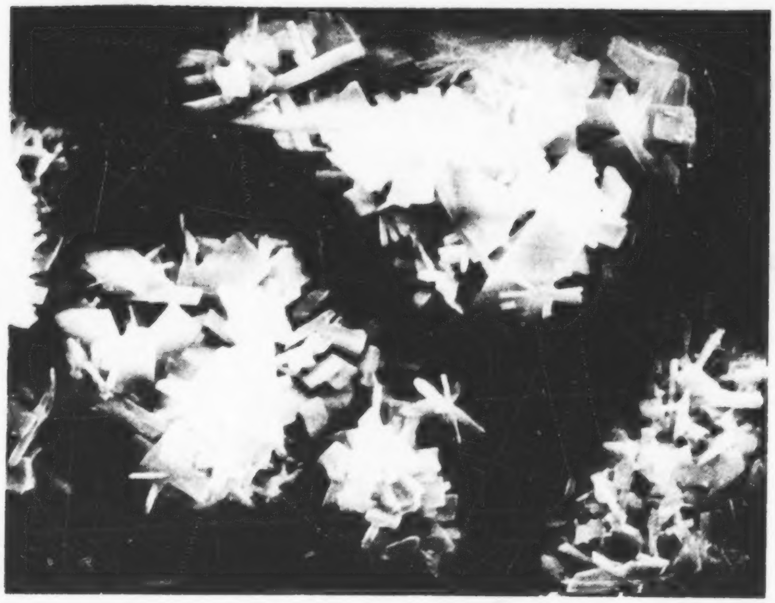
Four grinds of Aragonite limestone were tested; the principal results of these tests are listed in Table 5. At an approximately equivalent grind, pilot plant data indicated that Aragonite is much more reactive than Sylacauga and results in better performance. For example, with a nominal grind of 90% less than 200 mesh, run AR-3 showed that the pH of regenerated liquor reached 6.8 with 1.0 limestone stoichiometry of Aragonite. On the other hand, run SY-4 of Table 4 showed that a liquor pH of only 6.6 was attained with Sylacauga at 1.11 stoichiometry. Phenomenal solids dewatering was obtained with Aragonite: with 90% less than 325 mesh, the filter cake insoluble solids concentration reached 66%. Even with 90% less than 200 mesh, the filter cake solids content attained 62%. Only 43 to 48% solids in the filter cakes were obtained from Sylacauga with those two grinds. Pictures taken by SEM, as shown in Figures 3 and 4, show that solid products obtained from Aragonite and Sylacauga have essentially the same crystal morphology -- clusters of platelets. However, the platelets generated from Aragonite are much larger and thicker than those from Sylacauga.

Effects of limestone grind on liquor pH, SO_2 removal efficiency, and solids qualities obtained from Aragonite had the same trends as those from Sylacauga. At constant limestone feed rate, the coarser grind Aragonite resulted in lower pH, SO_2 removal efficiency, and filter cake insoluble solids.

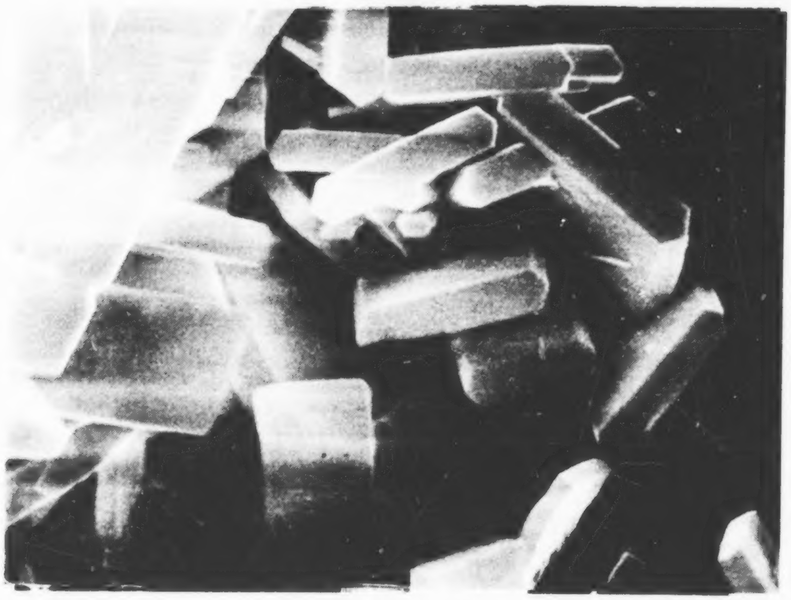
TABLE 5. RESULTS OF DUAL ALKALI PILOT TESTS WITH ARAGONITE

Run No.	AR-1	AR-2	AR-3	AR-4
Nominal grind, mesh	90%-325	90%-140	90%-200	90%-80
Scrubber feed pH	6.7	6.5	6.8	6.2
Scrubber effluent pH	6.3	6.1	6.4	5.7
Inlet SO_2, ppm	3010	3020	3060	3040
SO_2 feed rate, kg/hr	2.9	2.9	2.9	2.9
SO_2 absorption, %	94	90	94	83
Limestone slurry feed rate, kg/hr	9.5	9.5	9.5	9.5
Flue gas O_2, %	8.5	9.1	9.5	8.2
Filter cake solids, % (insoluble)	66	47	62	43
Total alkalinity, gmol/L	0.4	0.36	0.42	0.27
Run time, hr	72	86	88	52
Limestone stoichiometry	1.0	1.05	1.0	1.14
Settling rate (reactor), cm/min.	2.2	2.6	2.9	2.7
Settling test % solids	1.6	2.2	1.9	1.8
Active Na, gmol/L	1.02	1.04	1.02	0.93
SO_4/SO_3	1.9	2.1	1.8	2.7

Reproduced from
best available copy.

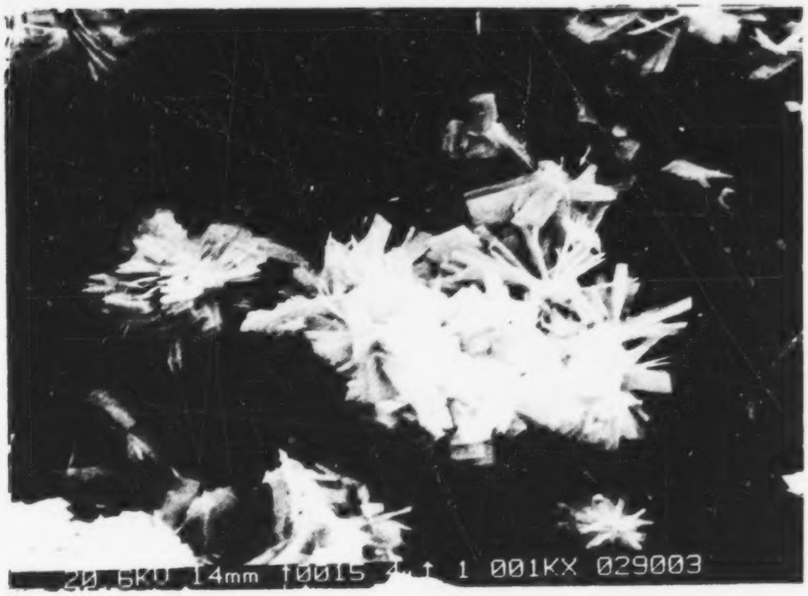


(a) 1000X

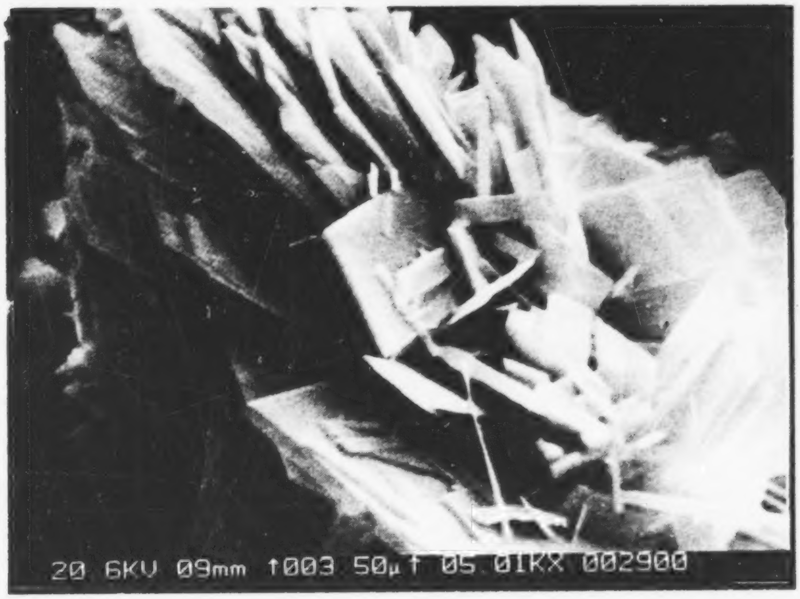


(b) 5000X

Figure 3. Scanning electron micrographs of solid products from Aragonite (run AR-1, 90% less than 325 mesh).


20.6KV 14mm ↑0015 4 ↑ 1 001KX 029003

(a) 1000X


20 6KV 09mm ↑003 50µ↑ 05 01KX 002900

(b) 5000X

Figure 4. Scanning electron micrographs of solid products from Sylacauga (run SY-1, 99% less than 325 mesh).

The only operating problem encountered during the Aragonite testing was foaming and ensuing losses of system liquor and unreacted limestone. For example, when testing with 90% less than 200 mesh Aragonite, considerable foaming occurred throughout the system, especially in the first reactor and the scrubber bleed hold tank. The foamy bubbles rose above the reactor top and overflowed, causing significant liquor loss. Since limestone was fed to the first reactor, the overflow also caused losses of unreacted limestone. As a result, the pH of regenerated liquor dropped to 6.5, and SO_2 removal efficiency decreased to 92%.* The foam was probably caused by oily organic matter naturally present as an impurity in the Aragonite. The organic matter also caused a high total carbonate concentration (about 5 g/L as CO_3) in the system liquor.

In order to solve the foaming problem, a defoamer, AZS Products D-103, was used. The defoamer is described as "esters in an alcohol base" and has a brownish color and characteristic smell of esters. Laboratory tests indicated that only a few parts per million (ppm) are needed to control foaming. The defoamer was mixed with limestone slurry and fed to the first reactor continuously. A mass balance indicated that the consumption rate of D-103 defoamer was about 0.125 g/kg (1/4 lb/ton) of limestone.

Once the foaming problem was under control, the pH of regenerated liquor rose to 6.8, and SO_2 removal efficiency increased to 94% with limestone stoichiometry at 1.0.

Fredonia

Five runs were made to test three grinds of Fredonia limestone. The results shown in Table 6 indicated that, at approximately equivalent grind,

---

*Data were not included in Table 5.

TABLE 6. RESULTS OF DUAL ALKALI PILOT TESTS WITH FREDONIA

Run No.	FR-1	FR-2	FR-3	FR-4	FR-5
Nominal grind, mesh	99%-325	90%-325	90%-325	90%-200	90%-200
Scrubber feed pH	6.7	6.6	6.7	6.5	6.7
Scrubber effluent pH	6.3	6.2	6.2	6.1	6.3
Inlet SO_2, ppm	3020	3010	3020	3040	3030
SO_2 feed rate, kg/hr	2.9	2.9	2.9	2.9	2.9
SO_2 absorption, %	93	92	93	91	92
Limestone slurry feed rate, kg/hr	9.5	9.5	10.5	9.5	11.4
Flue gas O_2, %	7.2	7.7	7.6	7.6	7.6
Filter cake solids, % (insoluble)	58	52	53	48	48
Total alkalinity, gmol/L	0.39	0.39	0.40	0.38	0.40
Run time, hr	82	88	72	88	65
Limestone stoichiometry	1.0	1.01	1.1	1.02	1.2
Settling rate (reactor), cm/min.	2.2	2.1	2.2	1.9	1.8
Settling test % solids	1.9	2.0	2.0	2.5	2.6
Active Na, gmol/L	1.07	1.07	1.09	1.07	1.1
SO_4/SO_3	2.0	2.0	2.0	2.2	2.1

Fredonia's performance was lower than that of Aragonite but higher than that of Sylacauga. For example, at a constant limestone feed rate, 90% less than 325 mesh Fredonia resulted in a 52% solids filter cake, lower than the 66% with Aragonite but higher than the 48% with Sylacauga. Similar relative performance for the three types of limestones was also obtained with respect to liquor pH and SO_2 removal efficiency.

Distinct limestone grind effects were also obtained with Fredonia. Finer grind always gave better performance. After experiencing a pH drop from 6.7 (run FR-1, fine grind Fredonia) to 6.6 (run FR-2, coarse grind Fredonia), the limestone feed rate of coarse Fredonia was raised to increase the liquor pH to 6.7 (run FR-3). Adding about 10% more limestone, the pH of regenerated liquor reached 6.7, which also resulted in slight increases of SO_2 removal efficiency. However, little change in filter cake dewatering properties was obtained by the addition of extra limestone. When similar actions were taken with the coarser ground Fredonia (run FR-5); more limestone (about 20 percent) was needed to raise the pH to 6.7. However, filter cake insoluble solids remained the same (48%) as run FR-4.

Tate

Four grinds of Tate limestone were available for pilot plant evaluation. However, due to the operational problems, only three grinds were tested (Table 7).

The pilot plant testing started with the finest grind (run TA-1, 99% less than 325 mesh). Due to the moderately high magnesium content in Tate limestone, significant accumulation of Mg^{2+} was obtained in scrubbing liquor. The test was started with scrubbing liquor containing 270 ppm Mg^{2+}. After 82 hours of operation, the Mg^{2+} concentration reached 680 ppm. Filter cake analyses also indicated 4.1 mg magnesium/g dry cake, considerably higher than

TABLE 7. RESULTS OF DUAL ALKALI PILOT TESTS WITH TATE

Run No.	TA-1	TA-2	TA-3	TA-4
Nominal grind, mesh	99%-325	99%-325	90%-325	90%-200
Scrubber feed pH	6.4	6.3	6.3	6.2
Scrubber effluent pH	6.0	5.9	5.9	5.8
Inlet SO_2, ppm	3030	3040	3010	3050
SO_2 feed rate, kg/hr	2.9	2.9	2.9	2.9
SO_2 absorption, %	90	90	87	85
Limestone slurry feed rate, kg/hr	9.5	9.5	9.5	9.5
Flue gas O_2, %	7.2	7.1	7.3	8.1
Filter cake solids, % (insoluble)	47	41	34	31
Total alkalinity, gmol/L	0.39	0.36	0.32	0.26
Run time, hr	82	46	86	38
Limestone stoichiometry	1.03	1.04	1.09	1.11
Settling rate (reactor), cm/min.	2.5	2.4	1.9	0.4
Settling test % solids	2.0	2.1	2.5	4.0
Active Na, gmol/L	1.22	1.11	0.95	0.87
SO_4/SO_3	2.5	2.11	2.2	2.8

the 1 mg/g found from runs with low magnesium limestones (Fredonia, Sylacauga, and Aragonite).

Testing with 99% less than 325 mesh Tate continued with run TA-2. The magnesium ion concentration in the scrubbing liquor kept increasing, reaching 990 ppm toward the end of the run. Relatively poor overall system performance was obtained; i.e., the pH of regenerated liquor was 6.3, the SO_2 removal efficiency was 90%, and the filter cake insoluble solids were only 41%.

Run TA-3 used 90% less than 325 mesh Tate. The magnesium ion concentration increased slightly to 1,120 ppm. However, the filter cake insoluble solids decreased to 34%. In addition, the pH of regenerated liquor remained at 6.3, but only 87% SO_2 removal efficiency was obtained.

The evaluation tests with Tate limestone were concluded with 90% less than 200 mesh grind (run TA-4). The pH of regenerated liquor dropped to 6.2, and SO_2 removal efficiency also dropped to 85%. A mass balance indicated that the limestone stoichiometry should be 1.11. Due to the low pH, the total alkalinity concentration decreased to 0.26 M and sulfate-to-sulfite ratio increased to 2.8. The filter cake insoluble solids concentration was only 31%. The slurry settling rate decreased considerably from 1.9 cm/min to 0.4 cm/min. After 34 hours of operation, system upset occurred and nonsettling solids overflowed from the thickener top to the forward feed hold tank. Due to the scaling and plugging problems and high insoluble solids concentration (greater than 3 percent) in the scrubber, the pilot plant was shut down and the nonsettling solids removed. The Mg^{2+} concentration had reached 1,180 ppm before shutdown.

Whitestone

Very poor scrubber performance was obtained following the use of high magnesium Whitestone (Table 8). The test series started with the coarse grind (run WH-1, 90% less than 200 mesh) and proceeded to finer grind. However, the

TABLE 8. RESULTS OF DUAL ALKALI PILOT TESTS WITH WHITESTONE

Run No.	WH-1	WH-2	WH-3	WH-4
Nominal grind, mesh	90%-200	90%-200	90%-325	90%-325
Scrubber feed pH	6.2	6.1	6.0	5.9
Scrubber effluent pH	5.7	5.7	5.5	5.2
Inlet SO_2, ppm	3070	3020	3030	3020
SO_2 feed rate, kg/hr	2.9	2.9	2.9	2.9
SO_2 absorption, %	85	84	83	76
Limestone slurry feed rate, kg/hr	10.0	10.9	11.4	10.0
Flue gas O_2, %	4.5	2.1	4.7	8.1
Filter cake solids, % (insoluble)	34	32	30	30
Total alkalinity, gmol/L	0.29	0.28	0.18	0.15
Run time, hr	72	56	74	38
Limestone stoichiometry	1.16	1.29	1.35	1.30
Settling rate (reactor), cm/min.	2.6	2.6	1.7	0.1
Settling test % solids	2.1	2.0	2.3	2.4
Active Na, gmol/L	1.16	1.10	0.98	0.93
SO_4/SO_3	2.2	2.3	3.5	3.9

negative effects of accumulated Mg^{2+} outweighed the beneficial effects of the finer limestone grind. Poorer performance was obtained as the Mg^{2+} concentration increased even though the reagent was switched to the finer ground Whitestone. The concentration of Mg^{2+} increased steadily from 590 to 1,510 ppm during this test series. The pH of regenerated liquor dropped to 5.9. At such a low pH, SO_2 evolution from the reactors and hold tanks caused serious concerns. The SO_2 removal efficiency also decreased significantly to about 76%. Filter cake insoluble solids never exceeded 34% and barely reached 30% toward the end of this testing series. System upset with nonsettling solids carried over in thickener overflow was encountered, causing control and scaling problems. Filter cake analyses confirmed that the limestone stoichiometry was 1.3. Cake analyses also showed greater than 10 mg magnesium/g dry cake.

Results from the Tate and Whitestone limestone testing confirmed previous findings by Chang et al.(7) that the limestone dual alkali system performance is a strong function of Mg^{2+} concentration through its effects on limestone dissolution rates. The effects of Mg^{2+} on the dissolution rate of limestone were interpreted by a surface adsorption model.(8) The model assumes that, due to the similarity between Ca^{2+} and Mg^{2+}, the Mg^{2+} can be adsorbed on the surface of $CaCO_3$ particles to form a Ca-Mg-carbonate surface. This adsorption reduces the limestone dissolution rate since the surface is partially blinded by adsorbed Mg^{2+}. Furthermore, if the surface adsorption of Mg^{2+} also occurs on $CaSO_3$ crystals, the adsorbed Mg^{2+} can act as an impurity which causes crystal defects and inhibits crystal growth. As a result, the properties of the solid product deteriorate with the increase of Mg^{2+} concentration as reflected by the decrease of slurry settling rate and filter cake insoluble solids. Detailed discussions of Mg^{2+} effects and surface adsorption model can be found in Reference 7.

LIMESTONE REACTIVITY CHARACTERIZATION

Pilot plant testing under this project demonstrated that the selection of appropriate type and grind of limestone has tremendous impacts on dual alkali system performance. Poor limestone can even result in chemistry upset and system shutdown. To preclude selection of a poor limestone, a means to quantitatively analyze the relative reactivities of various limestones is needed. The urgency of this need is illustrated by the diversely different reactivities discovered with the three high calcium limestones tested, and the inability to differentiate between them by their physical/chemical properties. The reactivity differences were so significant that better system performance was obtained with coarsely ground (90% less than 200 mesh) Aragonite than with finely ground (99% less than 325 mesh) Sylacauga. None of the chemical/physical analyses discussed (wet screen, BET, SEM, and chemical composition) can fully explain or predict the differences.

In order to facilitate the selection of limestone for full-scale dual alkali processes, titration techniques which have the potential of quantitatively defining limestone reactivities were evaluated. The apparatus used for the reactivity tests is shown in Figure 5. A beaker with an overflow port maintaining 200 ml of slurry was used to simulate the continuously stirred tank reactor of the dual alkali pilot plant. Limestone slurry with 5% solids was fed at a constant rate (about 10 ml/min) to the reactor. The pH of reactor slurry was maintained at 6.0 by titrating with acid solution. Limestone reactivity was defined as the quantity (ml) of acid added per ml of limestone slurry flowing through the reactor at 6.0 pH.

All the limestones tested in the pilot plant were evaluated in the bench-scale flow reactor. Solutions of hydrochloric acid (0.1 N) were first used as the titrant. The data scatter was high, especially when the high magnesium

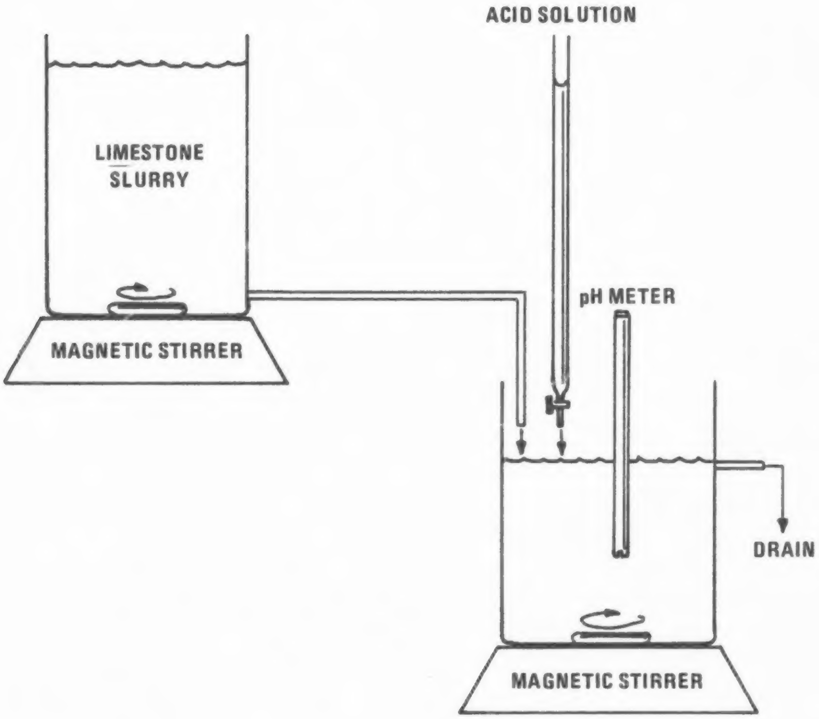

ACID SOLUTION
LIMESTONE SLURRY
MAGNETIC STIRRER
pH METER
DRAIN
MAGNETIC STIRRER

Figure 5. Apparatus for limestone reactivity laboratory tests.

limestone was titrated. Equivalent reactivities were obtained with Whitestone, Sylacauga, and Aragonite. However, pilot plant data indicated that Aragonite should be more reactive than Sylacauga, and that the system was not operable with Whitestone.

It was decided to use sodium bisulfite solution (0.4 N) to replace hydrocloric acid for limestone reactivity titration in an effort to:

- simulate the operating conditions in dual alkali reactors, and
- reduce the Ca^{2+} concentration by $CaSO_3$ precipitation.

Reagent grade sodium bisulfite was used to make the titrant. No unpleasant odor was detected, and no apparent SO_2 evolution was observed. New sodium bisulfite solutions were prepared each day to avoid oxidation. Sodium hydroxide and iodometric titrations were performed to confirm the acidity and bisulfite concentrations. The results of limestone reactivity tests are plotted against filter cake insoluble solids in Figure 6. Good correlations are shown in Figure 6 with not only low magnesium limestones (Sylacauga, Fredonia, and Aragonite), but also high magnesium ones (Whitestone and Tate). It was concluded that sodium bisulfite is preferable to hydrochloric acid in characterizing the limestone reactivities for dual alkali processes.

CONCLUSIONS

In summary, the pilot plant data indicate that:

- The type and grind of limestone has a significant impact on the performance and operability of dual alkali processes. Regenerated liquor pH, SO_2 removal efficiency, soda ash consumption, and filter cake dewatering properties are significantly affected by the limestone used.
- When maintaining constant limestone feed rate, the pH of regenerated liquor decreased with coarser grind. A drop in SO_2 removal efficiency

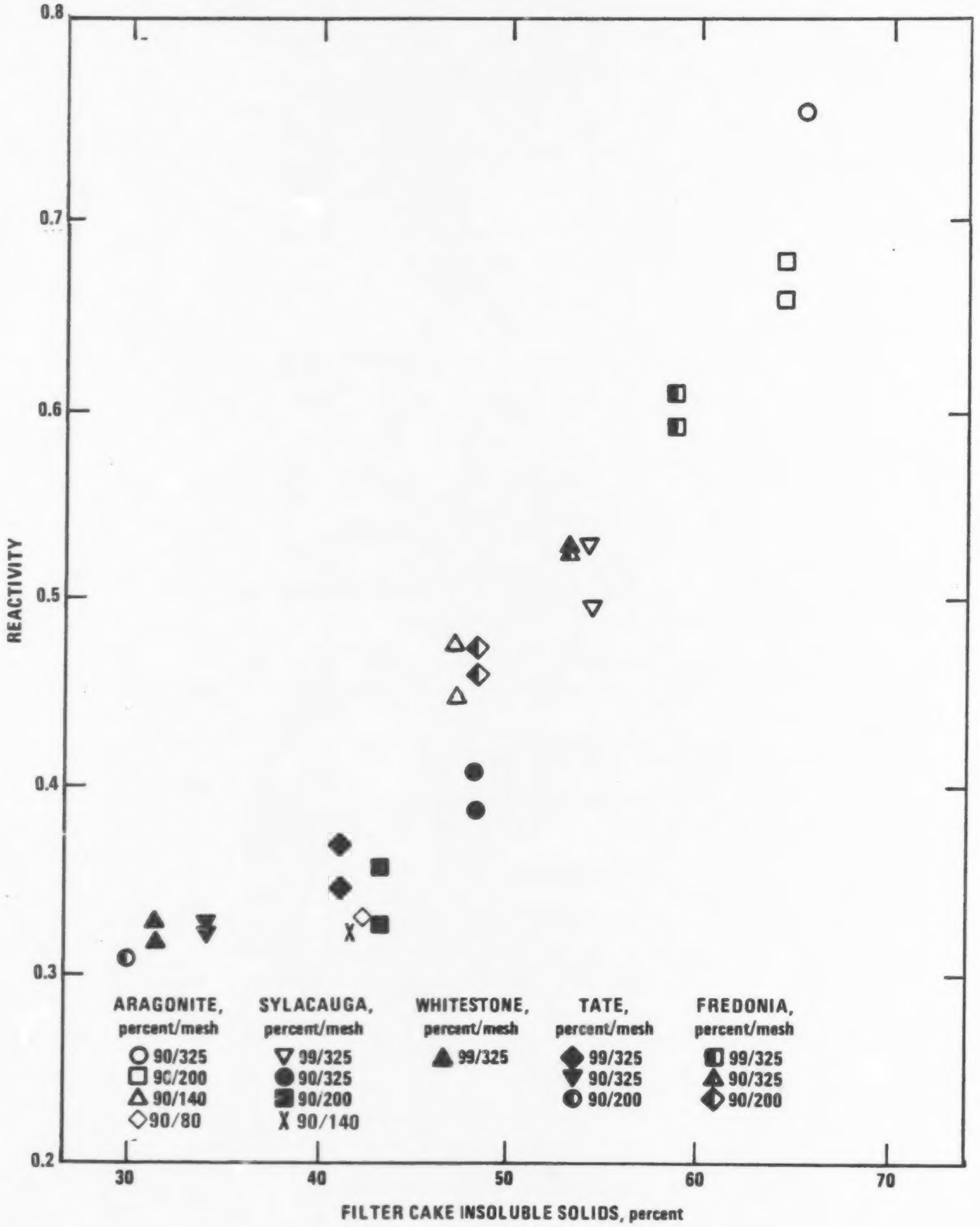


Figure 6. Limestone reactivity by $NaHSO_3$ titration vs. filter cake insoluble solids.

was also observed when the coarser grind limestone was used. The limestone stoichiometry had to be increased with coarser grinds to maintain constant pH and SO_2 removal. In addition to stoichiometry changes, the most significant result obtained was the effect of limestone grind on filter cake quality. Both slurry settling rate and filter cake insoluble solids concentrations decrease with the coarser grind limestones.

- Among the high calcium limestones tested, Aragonite is much more reactive than calcite. Superior performance was obtained with fine grind Aragonite, especially when considering solids dewatering properties. However, defoamers had to be used to avoid the foaming and ensuing system liquor losses with Aragonite.
- Poor system performance was obtained using high magnesium limestone. The scrubbing solution Mg^{2+} concentrations increased rapidly and exceeded 1,000 ppm within 150 hours of operation. The negative effects of accumulated Mg^{2+} outweighed the beneficial effects of the finer limestone grind. Poorer performance was obtained as the Mg^{2+} concentration increased even though the limestone was switched to finer grinds. System upsets were encountered with both high magnesium limestones tested.
- The data show that high purity, high calcium, fine ground limestone should be used for the regeneration of dual alkali scrubbing solution. Limestone containing more than 0.5% magnesium is not suitable.
- The relative reactivities of the limestones tested can be measured by a laboratory-scale titration method. Sodium bisulfite solution was used as the titrant to simulate the dual alkali reactors. Titration results correlate well with pilot plant data.

REFERENCES

1. Boynton, R. S., "Chemistry and Technology of Lime and Limestone," National Lime Association, John Wiley & Sons, New York, 1967.

2. Colley, J. D., O. W. Hargrove, and D. A. Stewart, "Effect of Limestone Grinding Circuit on FGD Performance and Economics," In Proceedings: Symposium on Flue Gas Desulfurization, New Orleans, LA, November 1983, Volume 1, EPA-600/9-84-017a (NTIS PB84-223031), July 1984.

3. Chang, C. S., J. H. Dempsey, and R. H. Borgwardt, "Effect of Limestone Type and Grind on SO_2 Scrubber Performance," Environmental Progress 1, 1, 59, February 1982.

4. Colley, J. D., R. L. Glover, T. Donaldson and D. A. Stewart, "Process Troubleshooting at a Utility Limestone FGD System," In Proceedings: Symposium on Flue Gas Desulfurization, New Orleans, LA, November 1983, Volume 1, EPA-600/9-84-017a (NTIS PB84-223031), July 1984.

5. Beals, J., L. Cannell, and J. Hengel, "How FGD Reagent Quality Affects System Performance," Power, 27, March 1984.

6. Chang, J. C. S. and N. Kaplan, "SO_2 Removal by Limestone Dual Alkali," Environmental Progress, 3, 4, 267, November 1984.

7. Chang, J. C. S., N. Kaplan, and T. G. Brna, "Effects of Magnesium and Chloride Ions on Limestone Dual Alkali System Performance," paper presented at the 189th national meeting of the American Chemical Society, Miami Beach, Florida, April 28 - May 3, 1985.

8. Sjoberg, E. L., "Kinetics and Mechanism of Calcite Dissolution in Aqueous Solutions at Low Temperatures, "Stockholm Contributions in Geology, 32(1), 1981.

Session 8: EMERGING TECHNOLOGIES

Richard G. Rhudy, Chairman
Electric Power Research Institute
Palo Alto, CA

PILOT EVALUATION OF COMBINED PARTICULATE AND SO_2 REMOVAL USING A FABRIC FILTER SYSTEM

Richard G. Hooper EPRI, Denver, CO
Verle Bland, KVB, Denver, CO
Franz G. Pohl, SoRI, Birmingham, AL
Michael McElroy and Richard Rhudy, EPRI, Palo Alto, CA

ABSTRACT

The process of combining SO_2 and particulate control using dry reagents is gaining increased utility interest. This paper reviews and updates events relevant to the injection of both sodium and calcium reagents into a flue-gas stream ahead of a fabric filter. Results are reported from tests conducted at EPRI's Fabric Filter Test Facility at the Arapahoe station of the Public Service Co. of Colorado in Denver, Colorado, and at EPRI's High-Sulfur Coal Pilot Plant at the Scholz station of the Gulf Power Company near Sneads, Florida.

With dry-sodium injection, sodium reagent is fed into the flue-gas stream ahead of a baghouse and downstream of the air heater. In the ductwork, sodium carbonate partially reacts with SO_2 in the flue gas to form a dry powder that collects along with the fly ash on the bags as part of the dustcake where it continues to remove SO_2. In this paper, results are presented for three sodium reagents -- sodium bicarbonate, sodium sesquicarbonate, and sodium carbonate. For each reagent, SO_2 removal is determined as a function of reagent particle size and normalized stoichiometric ratio (NSR), and sodium utilizations are compared.

Injection of a dry calcium reagent for SO_2 removal is performed in the same manner as sodium injection except that water or steam is also injected to condition the flue gas to a desired temperature and humidity. In the ductwork, a portion of the lime reacts with SO_2 to form calcium sulfite and calcium sulfate. In this paper, preliminary tests with pressure hydrated dolomitic lime are reported, and calcium utilization and SO_2 removal are compared for five calcium reagents -- quicklime, pressure hydrated high-calcium lime, conventionally hydrated lime, pressure hydrated dolomitic lime, and precalcined pressure hydrated dolomitic lime.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the view of the Agency and no official endorsement should be inferred.

INTRODUCTION

The Electric Power Research Institute (EPRI) is conducting tests at its Arapahoe Fabric Filter Pilot Plant, on low-sulfur coal, and at its Scholz High-Sulfur Coal Pilot Plant to characterize and assess combined SO_2 and particulate control for utility applications. This paper reviews and updates results relevant to flue-gas desulfurization (FGD) by the injection of dry sorbents upstream of a fabric filter. Dry-sodium injection is being tested at Arapahoe; dry-calcium injection is being tested at both Arapahoe and Scholz.

Dry-sorbent injection has the potential to greatly simplify FGD hardware requirements and to significantly reduce capital requirements for both new and retrofit SO_2 control applications. EPRI began studying dry-sodium injection in 1977 and to date has published results of a feasibility study (1), bench-scale investigation (2), and full-scale demonstration at the 22 MWe Cameo station of the Public Service Co. of Colorado (3). An economic evaluation of dry-sodium injection is scheduled to be published soon. More recently, investigations have begun on the possible use of calcium products as dry sorbents for flue-gas injection. This technology is in the very early stages of characterization and is not to be confused with in-furnace sorbent injection for combined NO_x-SO_x control, which is the subject of separate EPRI R&D projects. Preliminary research in dry-calcium injection was initiated by analyzing results from investigations in spray dryer technology where it was learned that a substantial portion of total SO_2 removal occurs in the baghouse. Studies using calcium were undertaken to determine if SO_2 removal comparable to that experienced using spray dryers could be achieved in system ductwork and baghouses without the need of expensive spray dryer vessels and slurry handling equipment.

With dry-sorbent injection, sodium or calcium reagent is fed into the flue-gas stream between the air heater and the baghouse. With dry-calcium injection, water or steam is introduced to regulate flue-gas temperature and relative humidity, in essence simulating conditions downstream of a spray dryer. In both cases (sodium and calcium), the reagent partially reacts with SO_2 in the flue gas and collects with fly ash on the bags in the baghouse as part of the dustcake, where it continues to remove SO_2. Both the spent reagent and the filtered fly ash are then removed in the normal course of bag cleaning and collected for disposal.

Figure 1 is a simplified flow schematic of a coal-fired power plant with dry-sorbent injection for SO_2 control. For both sodium and calcium, unit processes are much the same as those required for coal handling: transportation, storage, pulverization, and injection.

Injection of dry reagents for SO_2 control offers a number of advantages over alternate FGD options:

- Capital costs are significantly lower because of the comparative simplicity of the process.
- Only equipment already in common use at coal-fired power plants is required.

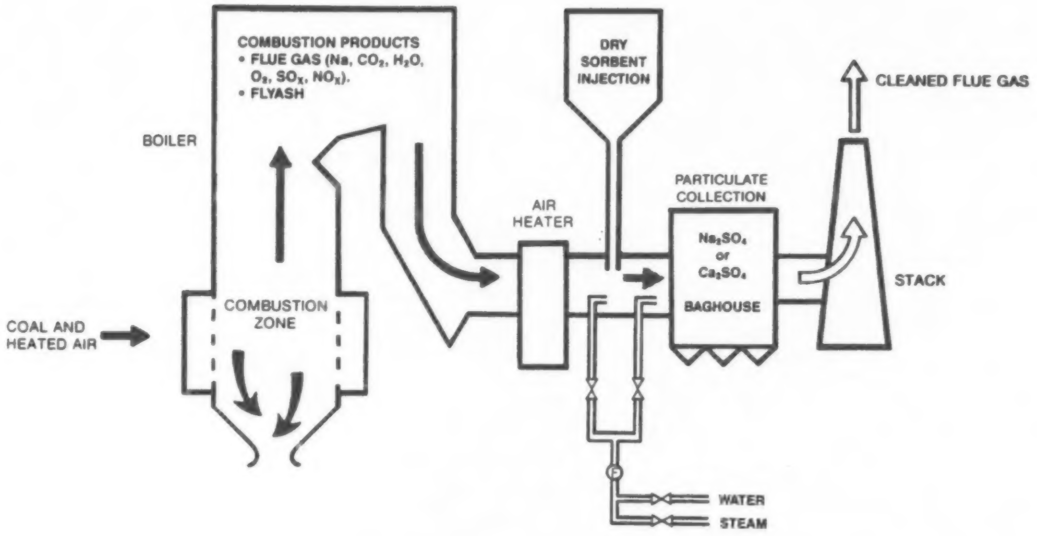

COMBUSTION PRODUCTS
• FLUE GAS (Na, CO₂, H₂O, O₂, SOx, NOx).
• FLYASH
BOILER
DRY SORBENT INJECTION
CLEANED FLUE GAS
AIR HEATER
PARTICULATE COLLECTION
Na₂SO₄ or Ca₂SO₄
BAGHOUSE
STACK
COMBUSTION ZONE
COAL AND HEATED AIR
WATER
STEAM

Figure 1. Schematic of a coal-fired power plant with dry-sorbent injection for SO_2 control.

- No wet slurry is required and no wet sludge is produced.
- The systems are retrofitted to boilers equipped with baghouses.
- Power costs are low.

DRY INJECTION OF SODIUM REAGENTS

With dry-sodium injection, sodium carbonate (Na_2CO_3) reacts with SO_2 to form sodium sulfate (Na_2SO_4), a dry powder. Most sodium reagents contain sodium bicarbonate ($NaHCO_3$), which decomposes to create sodium carbonate in a "popcorn" fashion, creating an open, porous microstructure that exposes more particle surface area. Typically, 20-30% of total SO_2 removal using sodium as the reagent occurs in the ductwork, and the remaining 70-80% in the baghouse.

Dry-sodium injection research to date at Arapahoe has involved three sodium reagents: sodium bicarbonate ($NaHCO_3$), sodium sesquicarbonate ($NaHCO_3 \cdot Na_2CO_3 \cdot 2H_2O$), and sodium carbonate (Na_2CO_3). For each reagent, SO_2 removal is determined as a function of reagent particle size and normalized stoichiometric ratio (NSR)*, and reagent utilizations are compared.

Description of Equipment

At the Arapahoe pilot plant, flue gas for dry sodium injection testing is withdrawn at the rate of 2200 acfm from Public Service Co. of Colorado's boiler number 4 (100 MW nominal load). Injection is accomplished by continuously moving sorbent from a small hopper via a screw feeder into the suction port of an eductor. Figure 2 illustrates the process. Approximately 5 cfm of carrier air transports the reagent materials into the duct at a point 6 feet upstream of the baghouse inlet. A scale weighs the feeder system during testing to determine reagent injection as a function of time.

Concentrations of SO_2, NO_x, and O_2 are continuously monitored upstream of the injection and after the baghouse, and relatively constant inlet SO_2 concentrations are maintained by an SO_2 dosing system. The pilot plant is in full-time operation using a three-hour reverse-gas bag cleaning cycle. Numerous controls and test parameters, including gas flow, temperature, and baghouse pressure drop are monitored throughout testing.

Discussion of Results

Sodium Bicarbonate-- Sodium bicarbonate was the first reagent tested at Arapahoe. Figure 3 presents SO_2 removal versus NSR for sodium bicarbonate of five different particle sizes: 11, 14, 32, 65, and 93 um.

---

*NSR is a measure of the quantity of sorbent injected relative to flue-gas SO_2 concentration and is defined as: $\frac{2 \text{ moles of Na}}{1 \text{ mole of S}}$. A NSR value of 1.0 results in two moles of sodium reacting with one mole of sulfur to produce one mole of Na_2SO_4 product.

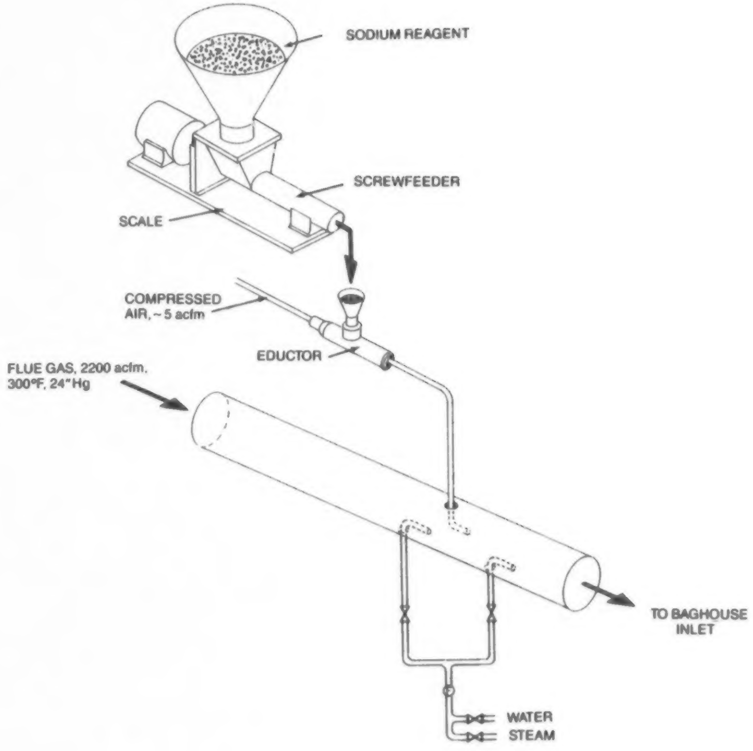

SODIUM REAGENT
SCREWFEEDER
SCALE
COMPRESSED AIR, ~5 acfm
EDUCTOR
FLUE GAS, 2200 acfm, 300°F, 24"Hg
TO BAGHOUSE INLET
WATER
STEAM

Figure 2. Illustration of dry-sodium injection at Arapahoe.

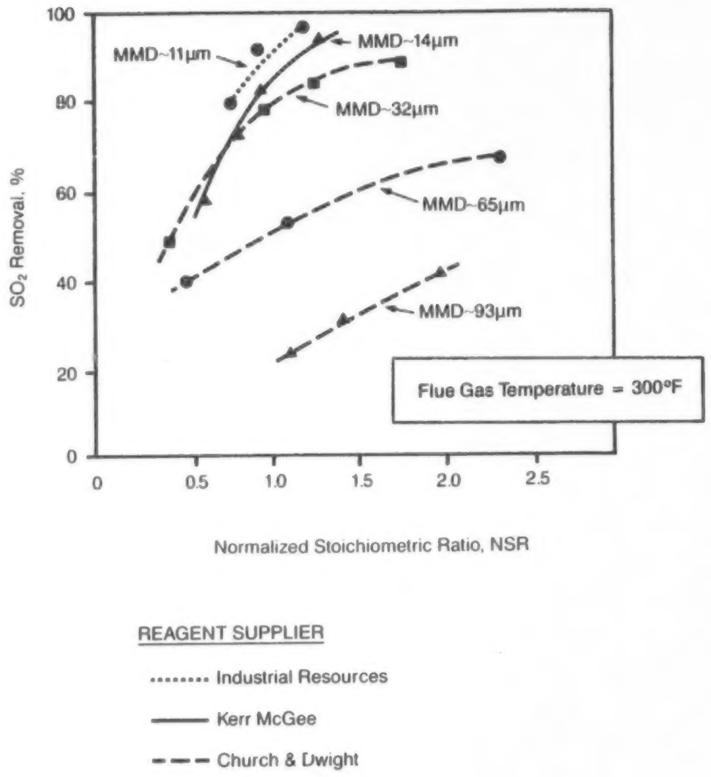


Figure 3. SO_2 removal vs. NSR for sodium bicarbonate illustrating the dependence of SO_2 removal on the MMD (mass mean diameter) of the reagent particles. Tests were conducted at Arapahoe on low-sulfur coal.

As can be seen in the figure, SO_2 removal is significantly greater for smaller reagent particle sizes, and removals greater than 90% are achieved with the finest particles. These data are consistent even though the reagents used were provided by three different suppliers -- Church and Dwight, Kerr-McGee, and Industrial Resources, Inc. -- and each supplier used a different method to mine and process its reagent.

The greater SO_2 removal for small reagent particles is not due to increased specific surface area. In fact, sodium bicarbonate of widely differing initial diameters decompose to particles of sodium carbonate with similar surface areas (about 10 times that of the original bicarbonate). Instead, preliminary analysis indicates that the greater removal is caused by better reagent distribution throughout the baghouse. Figure 4 is an illustration of the baghouse compartment at the Arapahoe pilot plant. As can be seen, the path required for particles to reach a bag thimble in one of the corners nearest the inlet duct is much less direct than that to reach a bag directly opposite the inlet. For maximum SO_2 removal, particles must be small enough to follow the gas stream and distribute evenly across the tube sheet to each bag. The result of maldistribution is that some bags become enriched with reagent while others are starved.

For these tests, the reagent was injected 6 feet upstream of the baghouse inlet. Flue gas velocity at the point of injection was approximately 60 fps, resulting in a sorbent in-duct residence time of less than 0.1 seconds. Future tests will be conducted to increase residence time to nearly one second. If this testing shows a significantly increased fraction of total SO_2 removed prior to collection on the bag, then in-duct residence time could be used to reduce the importance of reagent maldistribution across the tube sheet.

These findings are important when considering injection design criteria, reagent pulverization requirements and techniques, and possibly baghouse inlet transition and hopper designs. For example, the requirements on a pulverizer for dry injection might be slightly more demanding (i.e., finer particles required) when boiler load is reduced and gas flows are lessened. This underlines the importance of a good aerodynamic modeling effort relating to the injection system and baghouse designs, especially at lower boiler steam loads where flue-gas flow rates are substantially lower than design specifications.

Sodium Sesquicarbonate-- Four particle sizes of sodium sesquicarbonate were injected: 23, 28, 37, and 55 μm. Results are shown in Figure 5. As can be seen, the data trends are similar to those for sodium bicarbonate, with SO_2 removal again a strong function of reagent particle size. For this testing, both trona (naturally occurring sodium sesquicarbonate) and a commercial grade of synthetic, or processed, sodium sesquicarbonate were used.

Sodium Carbonate (Soda Ash)-- Most currently available soda ash compounds do not provide enough reaction surface area for efficient interaction with SO_2. As a result, once the original surface has reacted to form a layer of sodium sulfate, SO_2 cannot diffuse into the reagent to react further. Previously reported results from tests that directly inject soda ash into flue gas have

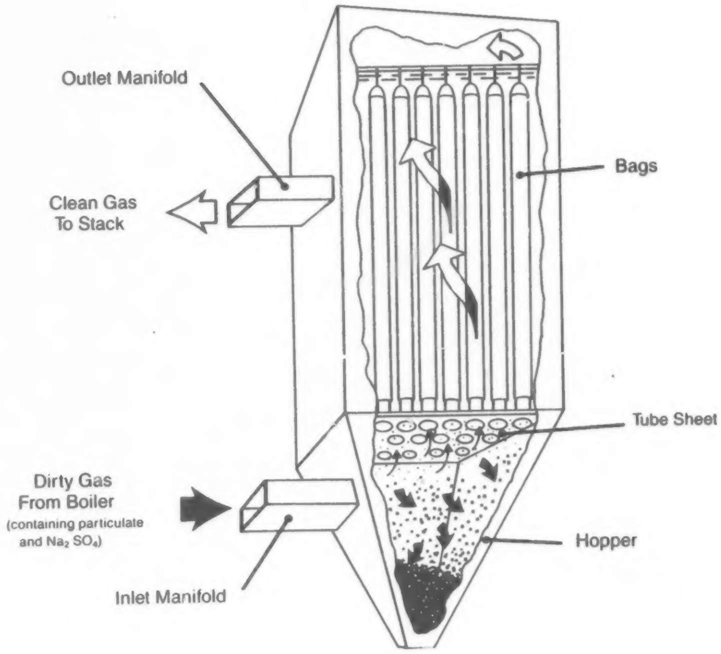

Outlet Manifold
Clean Gas To Stack
Bags
Tube Sheet
Dirty Gas From Boiler (containing particulate and Na₂ SO₄)
Hopper
Inlet Manifold

Figure 4. Baghouse compartment at the Arapahoe pilot plant.

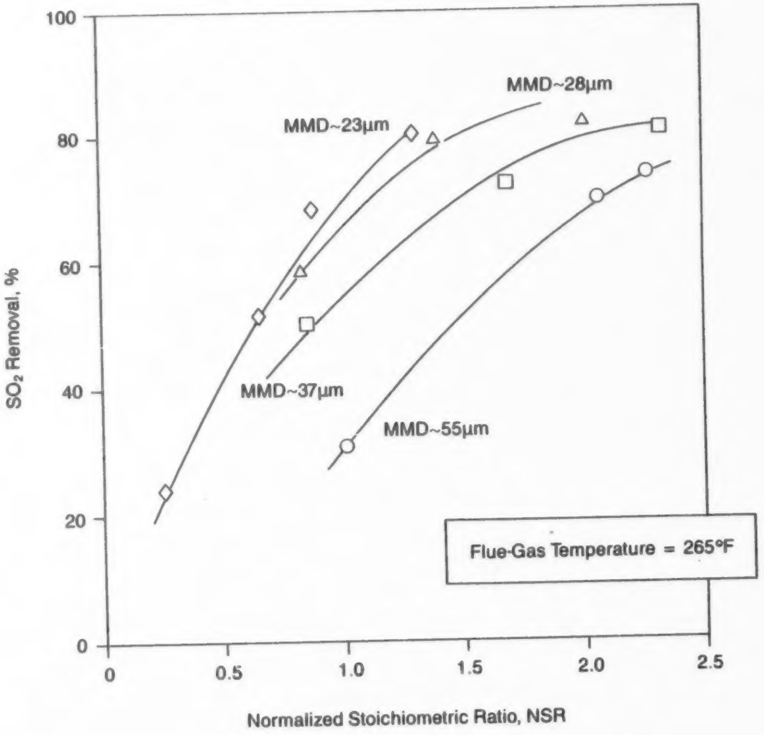
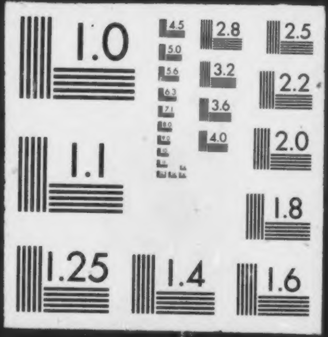


Figure 5. SO_2 removal vs. NSR for sodium sesquicarbonate illustrating the dependence of SO_2 removal on the MMD (mass mean diameter) of the reagent particles. Tests were conducted at Arapahoe on low-sulfur coal.

2 OF 7

PB86

138641


1.0
4.5
5.0
5.6
6.3
7.1
8.0
2.8
3.2
3.6
4.0
2.5
2.2
2.0
1.1
1.8
1.25
1.4
1.6

demonstrated SO_2 removals below 20% for NSR values greater than two. In an attempt to improve this performance, testing at Arapahoe was conducted with a "light" sodium carbonate provided by Kerr-McGee. This ash has a surface area of approximately 5 m^2/gm, significantly greater than the 1.5 m^2/gm (or less) of more conventional soda ash. Figure 6 shows test results at Arapahoe using this "light" sodium carbonate. As can be seen, for higher NSR values, SO_2 removal of almost 60% was achieved.

Sodium Utilization-- Results of an EPRI economic evaluation of dry-sodium sorbent injection emphasize that the delivered cost of the sodium is the most influential factor affecting long term system costs. It follows that sodium utilization can be critical in the economic feasibility of this technology.

Figure 7 presents sodium utilization as a function of NSR for three reagents with comparable initial particle diameters. In the figure, the highest sodium utilization values obtained by each reagent are plotted. For a given SO_2 removal, Figure 7 can be used to determine the NSR required and the utilization that would result for each reagent. To illustrate this, the curve corresponding to 70% SO_2 removal is included in the figure. As can be seen, to achieve 70% SO_2 removal, sodium bicarbonate displayed the greatest sodium utilization followed by sodium sesquicarbonate and then sodium carbonate.

Results presented in Figure 7 are for particles with MMDs of 15-30 μm. For particles that small, it can be assumed that reagent distribution is uniform across the tubesheets and consequently that differences in SO_2 removal between reagents are due to different final particle surface areas. Sodium bicarbonate "popcorns" into reactive sodium carbonate particles of much greater surface area, thereby increasing the probability of gas-solid contact and resulting in sodium utilization greater than 90%. The same effect occurs for the sesquicarbonate (composed of one molecule of sodium bicarbonate, one molecule of sodium carbonate, and two molecules of water), but because only part of the molecule can popcorn into sodium carbonate, there is a smaller proportional increase in surface area. As a result, sodium utilization for sodium sesquicarbonate was lower, ranging between 60-85%. Sodium carbonate has no popcorn effect and so its sodium utilization is directly related to the original surface area of the injected product. Utilization values for sodium carbonate were 25-35%. As a consequence of the popcorn effect, the trend of increasing SO_2 removal with decreasing initial reagent particle size is in general meaningful only for sodium reagents of the same composition.

DRY INJECTION OF CALCIUM REAGENTS

With dry-calcium injection, hydrated lime ($Ca(OH)_2$) reacts with SO_2 to form solid calcium sulfite ($CaSO_3$) and calcium sulfate ($CaSO_4$). This paper reports preliminary testing of dry-calcium injection using pressure hydrated dolomitic lime. In addition, SO_2 removal and calcium utilization are compared for five reagents -- quicklime, pressure hydrated high-calcium lime, conventionally hydrated lime, pressure hydrated dolomitic lime, and pre-calcined pressure hydrated dolomitic lime. This work was performed using high-sulfur coal at the Scholz pilot plant and low-sulfur coal at the Arapahoe pilot plant.

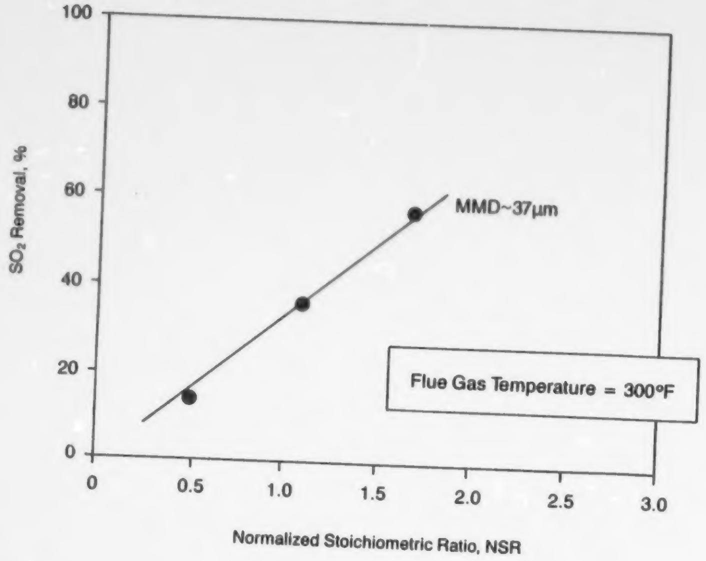


Figure 6. SO_2 removal vs. NSR for "light" sodium carbonate for one reagent particle size. Tests were conducted at Arapahoe on low-sulfur coal.

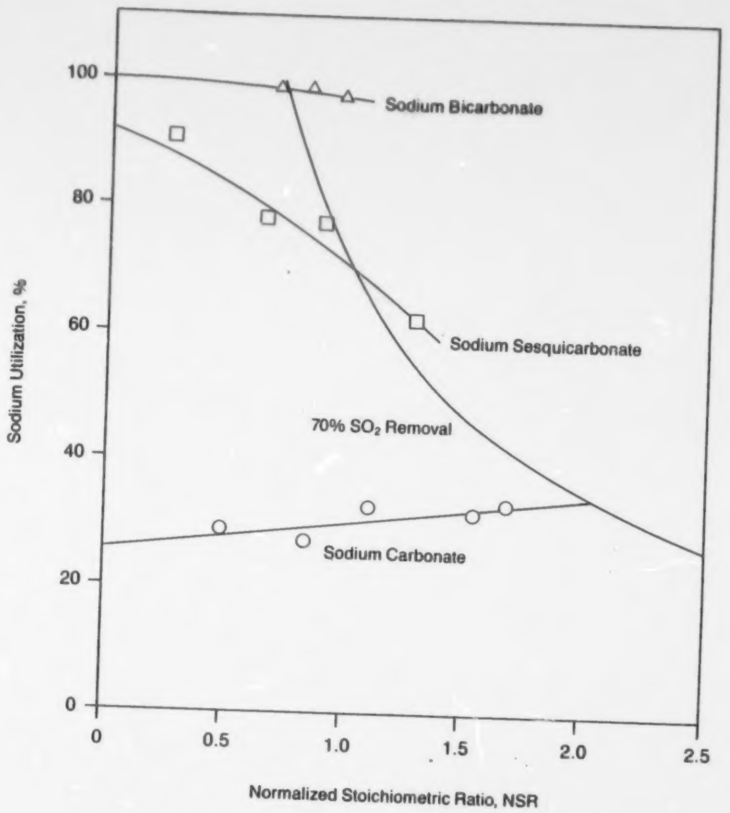


Figure 7. Sodium utilization vs. NSR for three sodium reagents tested at Arapahoe. Tests were conducted on low-sulfur coal.

Description of Equipment

The concept of injecting calcium compounds into low temperature flue gas is illustrated in Figure 8. Except for moisture injection, the basic hardware configuration is the same as for dry-sodium injection. In the testing, water and steam were injected both upstream and downstream of the reagent injection to evaluate the effect on operations of varying the location of the wet/dry interface.

Discussion of Results

Figure 9 presents results obtained at Scholz using high-sulfur coal. Although the data are scattered, the trend is towards higher SO_2 removal at lower approach-to-saturation temperatures. A contributor to the data scatter in the figure is that the tests involved were conducted at calcium-to-sulfur ratios (Ca/S) of 0.8-1.3.

Results from similar tests run using low-sulfur coal at the Arapahoe test facility are shown in Figure 10. Although preliminary, the data show that at flue-gas temperatures approaching saturation, dry injection of dolomitic lime can achieve 70-90% SO_2 removal. Also, the results confirm the trend at Scholz where SO_2 removal increased as flue-gas temperature decreased.

The relative sensitivity of SO_2 removal to the Ca/S ratio for pressure hydrated dolomitic lime is illustrated in both Figure 10 and Figure 11. Figure 11 also indicates that inlet SO_2 concentration does not strongly affect SO_2 removal. In the figure, most of the data were obtained at Arapahoe, where nominal inlet SO_2 concentrations were 300 ppm. However, a single point is shown from the results obtained at Scholz, where nominal concentrations were slightly greater than 2000 ppm. This single data point is included to illustrate that results obtained on both high- and low-sulfur coal are in good agreement.

Figure 12 presents SO_2 removal and sorbent utilization for five calcium reagents evaluated under similar test conditions. Each of these reagents was injected into flue gas conditioned to an approach-to-saturation temperature of about 25°F (flue-gas relative humidity varied between 45-69%). As can be seen in the figure, SO_2 removal varied between 23-78% and calcium utilization varied between 11-28%. In this comparison, the surface areas of the five materials differed considerably, which may have had a significant influence on the results. Although significant levels of SO_2 can be removed by dry injection of various calcium compounds, calcium utilizations are low. Additional dry-calcium injection testing is planned in 1985 to assess techniques for increasing SO_2 removal and improving calcium utilization.

CONCLUSIONS

Dry-sodium and dry-calcium sorbent injection offer a number of advantages over alternate FGD options, and research to better characterize and understand their application in utility settings is continuing. Development of dry-sodium injection technology is more advanced at this point in time. The issues now being investigated include sorbent particle size and sorbent utilization.

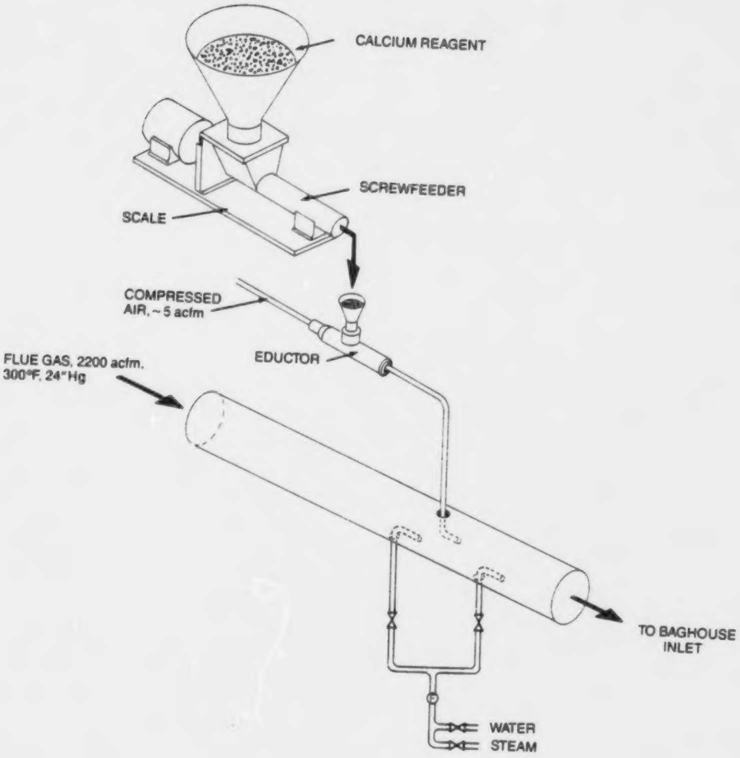

CALCIUM REAGENT
SCREWFEEDER
SCALE
COMPRESSED AIR, ~5 acfm
EDUCTOR
FLUE GAS, 2200 acfm, 300°F, 24"Hg
TO BAGHOUSE INLET
WATER
STEAM

Figure 8. Illustration of dry-calcium injection at Arapahoe and Scholz.

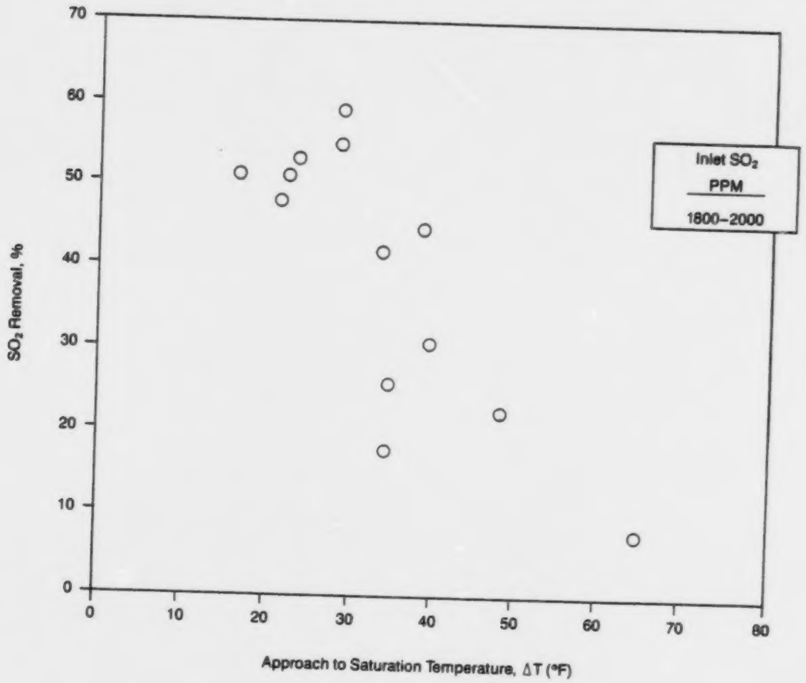


Figure 9. SO_2 removal vs. approach-to-saturation temperatures illustrating a data trend of higher SO_2 removal for lower approach-to-saturation temperatures. Results are for pressure hydrated dolomitic lime for a range of calcium-to-sulfur ratios of 0.8-1.3. Tests were conducted at Scholz on high-sulfur coal.

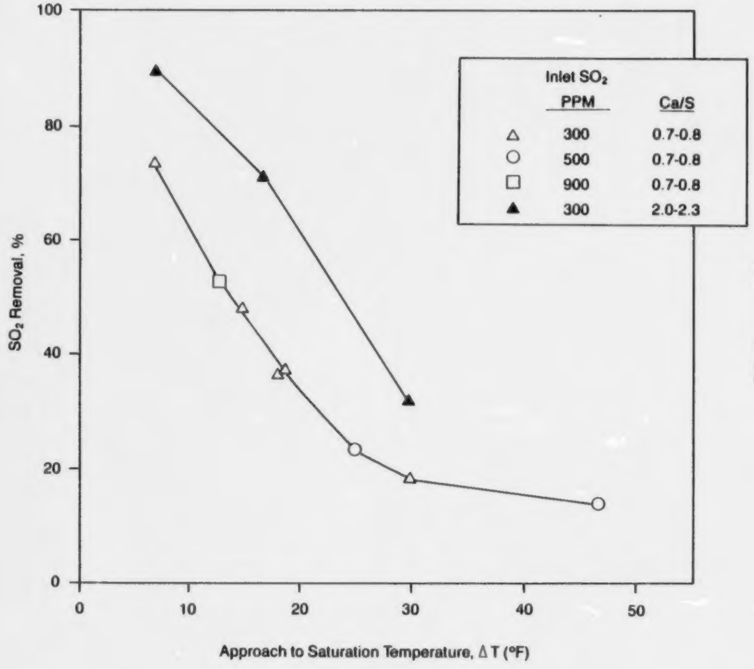


Figure 10. SO_2 removal vs. approach-to-saturation temperatures for pressure hydrated dolomitic lime illustrating the dependence of SO_2 removal on calcium-to-sulfur ratio. Tests were conducted at Arapahoe on low-sulfur coal.

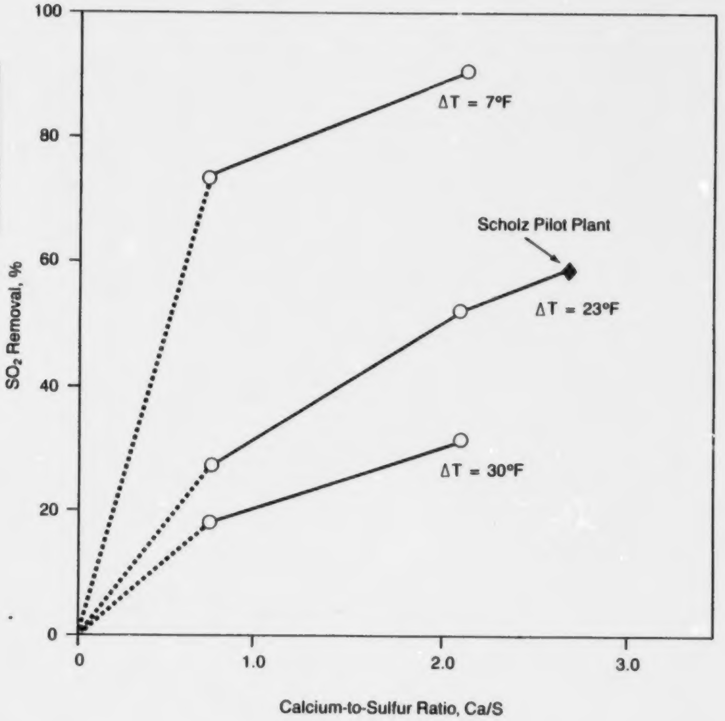


Figure 11. SO_2 removal vs. calcium-to-sulfur ratio for pressure hydrated dolomitic lime illustrating the dependence of SO_2 removal on approach-to-saturation temperature (ΔT). Tests were conducted at Arapahoe on low-sulfur coal (open circles) and at Scholz on high-sulfur coal (solid diamond)

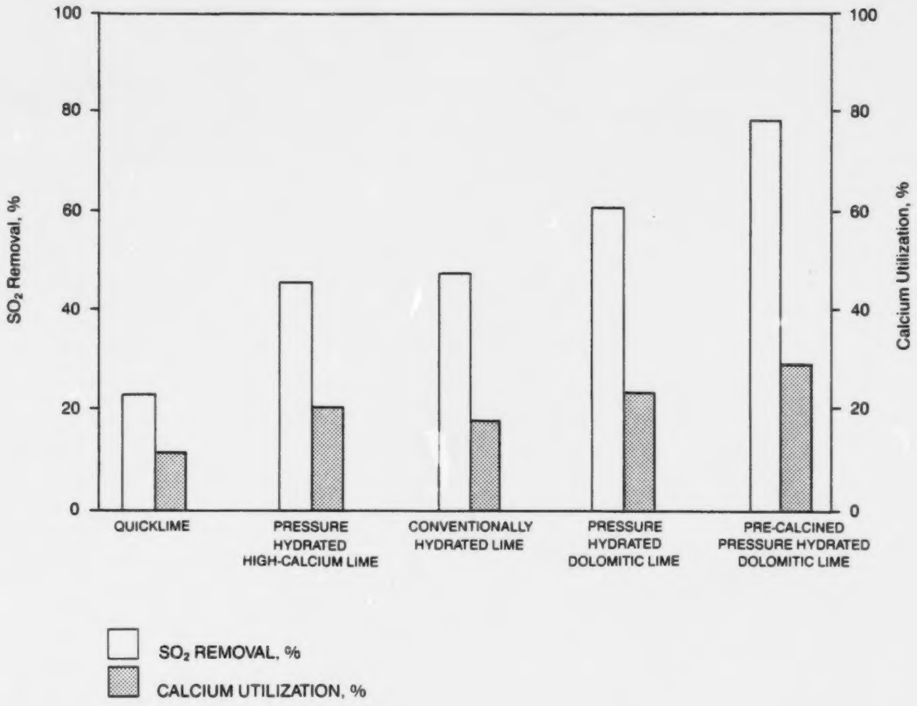


Figure 12. SO_2 removal and calcium utilization for five calcium reagents.

REFERENCES

1. Bechtel Corporation, "Evaluation of Dry Alkalis for Removing Sulfur Dioxide from Boiler Flue Gases," FP207, October, 1979, Electric Power Research Institute, Palo Alto, CA.

2. L.J. Muzio, J.K. Arand, "Bench-Scale Study of the Dry Removal of SO_2 with Nahcolite and Trona," CS-1744, March, 1984, Electric Power Research Institute, Palo Alto, CA.

3. L.J. Muzio, T.W. Sonnichsen, "Demonstration of SO_2 Removal on a 22 MW Coal-Fired Utility Boiler by Dry Injection of Nahcolite," CS-2894, Vol. 1, March, 1983; Vol. 2, June, 1984, Electric Power Research Institute, Palo Alto, CA.

INVESTIGATION OF COMBINED PARTICULATE AND SO_2 USING E-SOX

by: Leslie E. Sparks, Norman Plaks, Geddes H. Ramsey,
and Richard E. Valentine
Particulate Technology Branch
Air and Energy Engineering Research Laboratory
Environmental Protection Agency
Research Triangle Park, N. C. 27711

ABSTRACT

Research aimed at developing a low cost retrofit system, called E-SOX, for combined particulate and SO_2 removal is described. The E-SOX concept centers on recent advances in ESP technology that make it possible to reduce the size of ESPs required for particulate control. Results of mathematical modeling and limited pilot scale experiments on the feasibility of collecting SO_2 in the freed space are discussed. The results show that the concept is technically feasible. Modest (40-65%) levels of SO_2 removal are possible with lime as the reagent. High (over 90%) levels of SO_2 removal are possible with sodium carbonate as the reagent. Order of magnitude economic analysis shows that the process is economically feasible. Plans for additional experimental and theoretical work are presented.

INVESTIGATION OF COMBINED PARTICULATE AND SO_2 USING E-SOX

INTRODUCTION

Because of the concern over acid rain and the resulting interest in low cost methods for reducing sulfur dioxide (SO_2) emissions from existing plants, the Particulate Technology Branch has begun theoretical and experimental work to combine advanced ESP and SO_2 removal technologies in an existing ESP. This process, called E-SOX, combines advanced ESP technology and spray dryer technology to remove particulate and SO_2 in the same physical equipment.

Advanced ESP technology provides the means for collecting particles with high efficiency in relatively small ESPs. The technology can be used to reduce the required size of new ESPs or to free space in an existing ESP that can be used to collect SO_2.

This paper describes experiments and modeling which demonstrate the technical feasibility of E-SOX in retrofit situations.

CONCEPTUAL DESIGN

The basic idea of E-SOX is to turn the existing ESP into a combination spray dryer and advanced ESP. The conceptual design can best be understood by considering a hypothetical ESP used to control particulate emissions from a boiler firing high sulfur coal. Such an ESP would have a specific collector area (SCA) of 41.3 sec/m based on actual gas flow conditions with three 3 m long electrical sections in series. This size ESP is adequate to meet a particulate emission standard of 43 ng/J when collecting fly ash from a high sulfur coal.

The following modifications would convert the ESP to E-SOX:

1. Remove the wires and plates in electrical section 1.
2. Install a bank of spray nozzles at the inlet of electrical section 1. These nozzles should produce fine drops with diameters between 15 and 30 μm.
3. Install prechargers (either cooled pipe or other design) at the inlet of sections 2 and 3.
4. Replace the normal discharge electrodes in sections 2 and 3 with large diameter wires.

The removal of wires and plates in section 1 provides a space for SO_2 removal. At typical gas velocities in an ESP (1 to 2 m/sec), section 1 would provide about 1.5 to 3.0 sec residence time for SO_2 removal and drop evaporation.

The precharger and large diameter wires added to sections 2 and 3 convert the remainder of the ESP from single-to multi-stage. This conversion will compensate for the reduction in SCA from 41.3 to 27.5 sec/m. As is discussed later, this multi-stage ESP configuration makes it possible to meet the 43 ng/J particulate emission standard assumed for the plant.

The design goal for the E-SOX system is continued compliance with the particulate emission limits of 43 ng/J for the plant and 50 to 60% SO_2 removal. A particulate emission standard of 43 ng/J is typical of the particulate emission standards required of most existing power plants.

MODELING

E-SOX is designed to use part of the existing ESP as the reaction chamber for SO_2 removal. This means that all SO_2 removal and drying must occur in 1 to 2 sec residence time. If drying is not complete by this time, severe operating problems are likely in the active ESP sections following the reaction zones.

The spray dryer model developed by Damle (1) was used to determine the feasibility of collecting SO_2 in the 1 to 2 sec residence time available in the inlet section of an ESP. An example of the model projections is shown in Figure 1. The model predictions show that, if the droplet diameter is less than 30 μm, the solids entering the ESP should be dry and significant SO_2 removal is possible.

The modeling was favorable enough to warrant a limited experimental study aimed at verifying the model projections.

EXPERIMENTAL

The experiments were conducted in the AEERL in-house ESP pilot plant. The pilot ESP is a four-section single-lane pilot plant capable of operating from ambient to 350°C. Sampling ports are located at the ESP inlet, the ESP outlet, and between each pair of electrical sections of the ESP.

The configuration for these experiments is shown in Figure 2. Sections 1 and 2 were converted to SO_2 reaction chambers by removing the wires. The SO_2 was injected at point A located before the gas distribution plate. Scrubbing liquid was injected at point B using two sonic spray nozzles. These nozzles are two fluid nozzles and produce drops of 5 to 10 μm in diameter. SO_2 was measured at four locations (C, D, E, and F). Location C provided about 0.5 sec of residence time; location D, 1.3 sec; location E, 2.1 sec; and location F, 3.7 sec.

For the E-SOX tests the pilot plant was operated with a gas inlet temperature of 150°C and a gas flow rate of about 0.487 am^3/sec at 150°C. The ESP was operated under about 5 cm H_2O negative pressure to ensure that no SO_2 was released into the room. SO_2 and air flow rate were measured to determine the air leakage into the ESP. These measurements showed about 5% inleakage between the inlet and the sampling ports at the

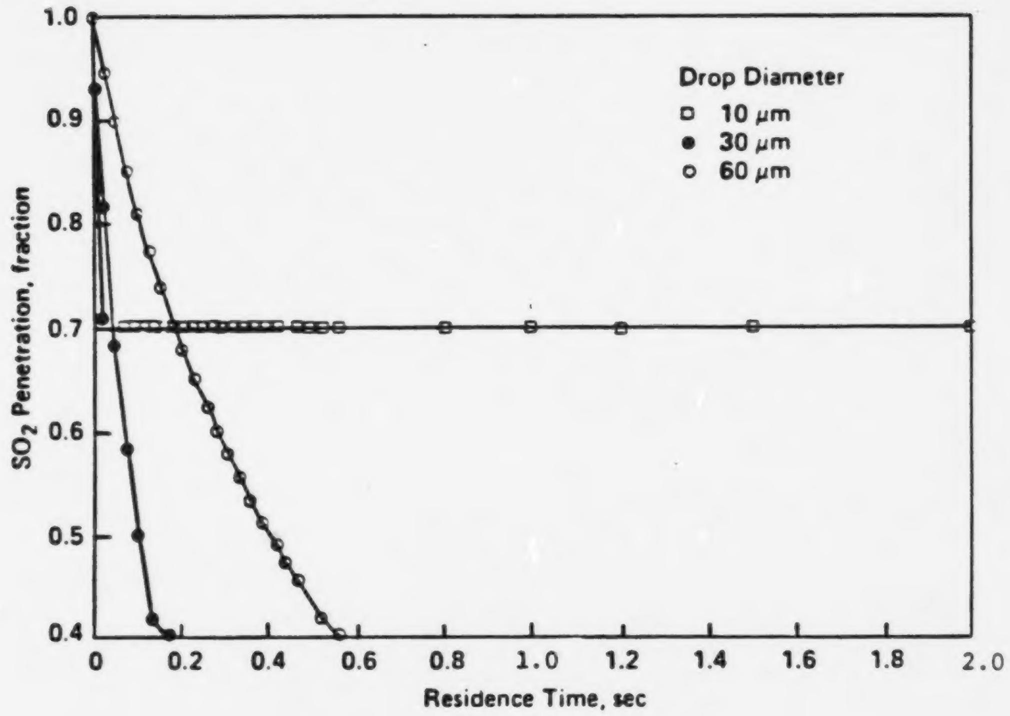


Figure 1. Predicted SO_2 removal for E-SOX.

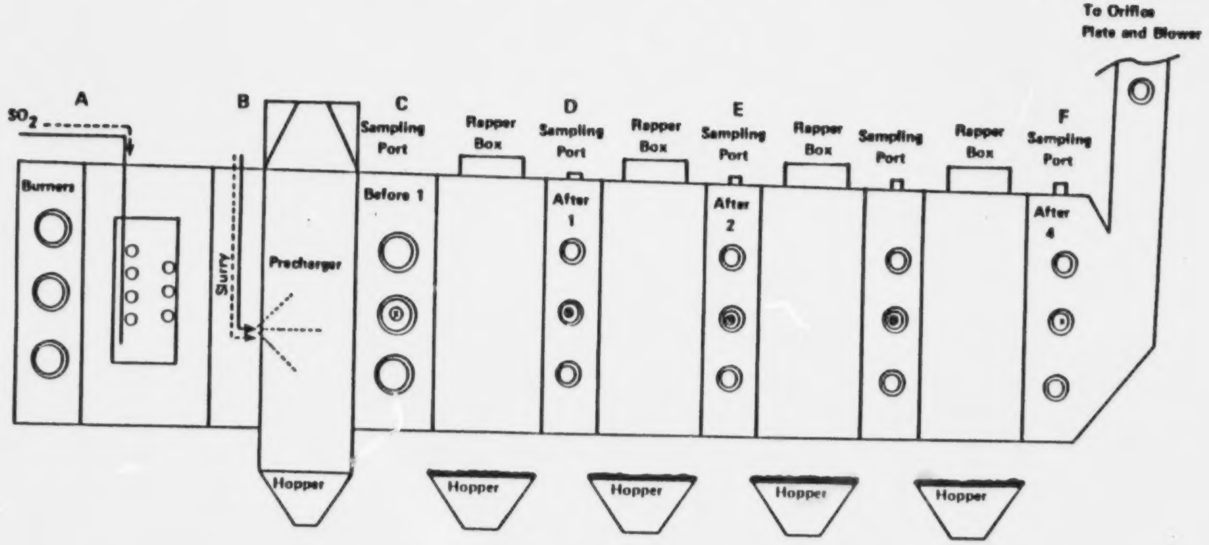

To Orifice
Plate and Blower
SO_2
A
B
C
Sampling
Port
Rapper
Box
D
Sampling
Port
Rapper
Box
E
Sampling
Port
Rapper
Box
Sampling
Port
Rapper
Box
F
Sampling
Port
Burners
Slurry
Precharger
Before 1
After
1
After
2
After
4
Hopper
Hopper
Hopper
Hopper
Hopper

Figure 2. Sketch of pilot plant modified for E-SOX

end of section 2. There was 20% inleakage between the ESP inlet and the ESP outlet. The E-SOX performance data were corrected for this inleakage.

SO_2 was measured with a TECO analyzer that was zeroed and spanned periodically. The SO_2 injection rate was between 1100 and 1500 ppm.

Preliminary measurements were made to determine the uniformity of the SO_2 concentration in the ESP. Horizontal traverses were made at three vertical locations at each sampling location. These measurements showed no vertical stratification of the SO_2, but a slight horizontal stratification. Because there was no vertical stratification, all experimental measurements were made using the center port at each sampling location. A multipoint horizontal traverse was made to account for the small amount of horizontal stratification.

Two sets of experiments were conducted: sodium carbonate was used as the scrubbing reagent for the first set; and lime was used for the second. The experimental procedure was the same for both sets of experiments. SO_2 readings were made at all sampling locations with the scrubbing liquid off. The scrubbing liquid was turned on and the system allowed to come to steady state. Then SO_2 measurements were made and the scrubbing liquid was turned off. The SO_2 measurements were then repeated. For a typical run the maximum difference between the two scrubbing-liquid-off measurements was less than 20 ppm; i.e., the inlet SO_2 concentration varied by less than 20 ppm.

RESULTS

Sodium Carbonate

Initial experiments were conducted using sodium carbonate solution as the scrubbing liquid. The results of these measurements are summarized in Figures 3 and 4.

The SO_2 concentration dropped rapidly during the initial 0.5 sec. The SO_2 concentration then dropped nearly linearly for the remainder of the time in the ESP. The initial rapid decrease in SO_2 is due to the reaction of SO_2 with the evaporating drops. During this phase the mass transfer of SO_2 to the drops is rapid, as is the reaction with the dissolved SO_2 and the sodium carbonate.

The linear phase is due to the reaction of the SO_2 with dry sodium carbonate particles. The gas-solid mass transfer is much less efficient than the gas-liquid mass transfer. However, the reactivity of sodium carbonate is high enough to ensure that considerable SO_2 is removed after the drops have evaporated.

The data in Figure 3 indicate that significant SO_2 removal is possible with E-SOX at reasonable stoichiometry. At high stoichiometry, over 90% removal can be realized.

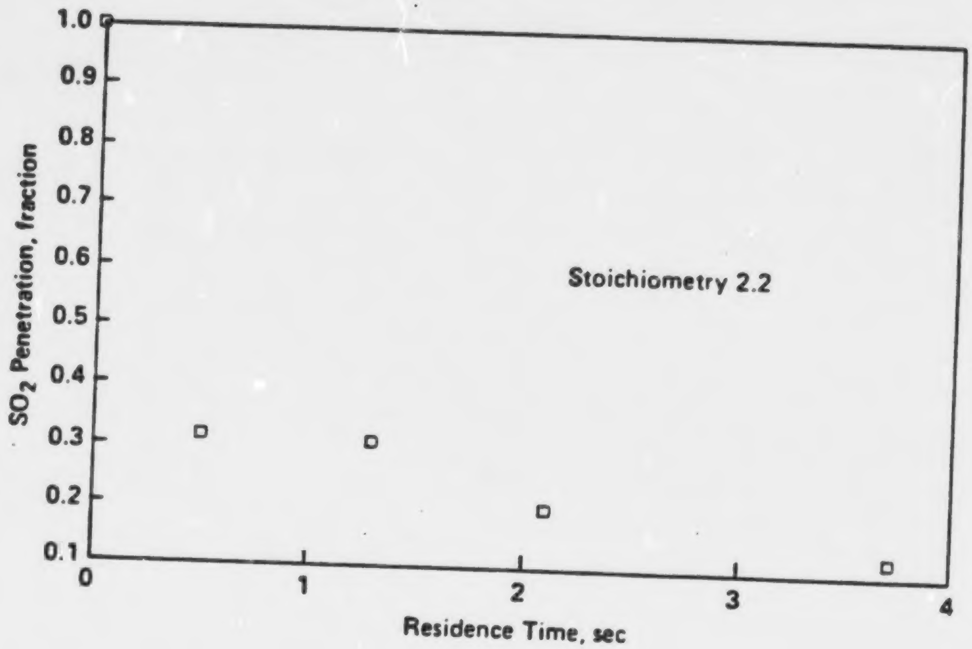


Figure 3. E-SOX SO_2 penetration vs. time.

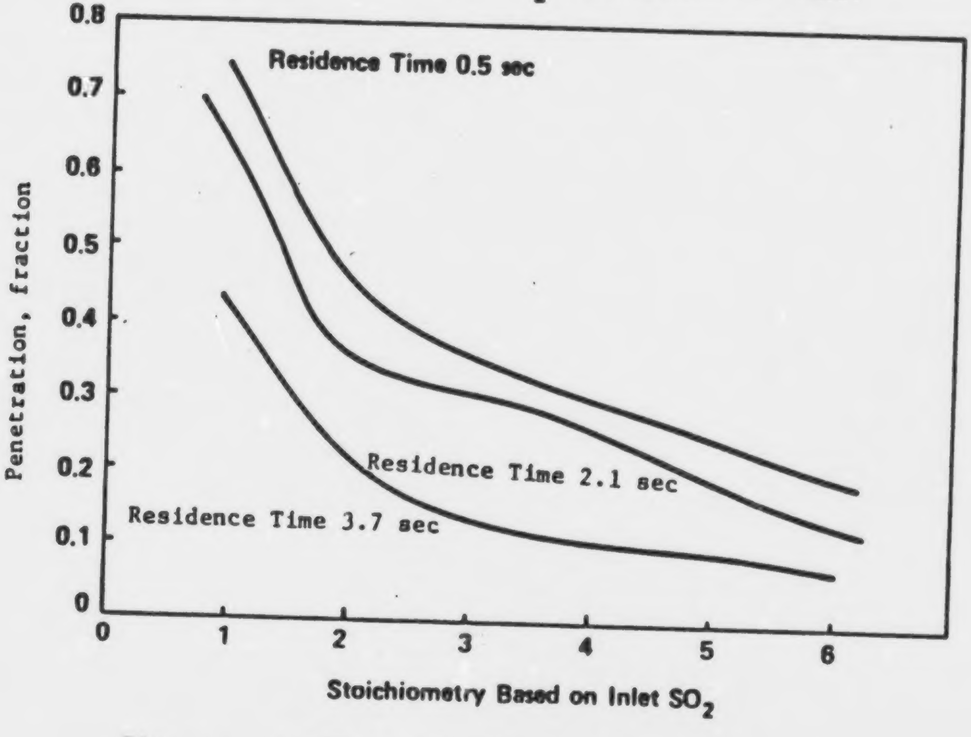


Figure 4. E-SOX experimental data for sodium carbonate

Lime

Because the results with sodium carbonate were encouraging, experiments were conducted using hydrated lime as the alkaline material. A slurry of lime and water was injected through the same nozzles used during the sodium carbonate experiments. The results of the lime experiments are shown in Figures 5 and 6.

The results in Figures 5 and 6 demonstrate that with lime all the SO_2 removal is obtained during the initial phase when SO_2 is collected by the evaporating drops. The dry lime used in these particles was not reactive enough to allow measurable SO_2 removal due to the dry lime particles. This indicates that the major advantage of sodium carbonate as compared to lime is the high reactivity of the dry sodium carbonate particles. Some of the highly reactive limes, being studied in conjunction with other retrofit SO_2 removal systems, may increase the SO_2 capture efficiency of E-SOX with lime to near the levels obtained with sodium carbonate.

The experimental data for both lime and sodium carbonate show that the design goals of the E-SOX system can be achieved with either reagent. A residence time of 1 sec is sufficient for 60% SO_2 removal with E-SOX and lime.

PARTICULATE COLLECTION EXPERIMENTS

E-SOX's success depends on its ability both to provide 50-60% SO_2 removal and to meet particulate emission limitations. Therefore, experiments were conducted to simulate the particulate collection of an E-SOX ESP. In these experiments lime slurry was injected through the spray nozzles. Fly ash was injected through sand blast guns, as is normal for the operation of the pilot plant for fly ash experiments. The ESP was converted to multi-stage configuration with cold-pipe prechargers installed in sections 1 and 3 and large wires installed in sections 2 and 4. The SCA of the ESP for these tests was about 11 sec/m.

Calibrated cascade impactors were used to determine the particle size distribution entering and leaving the multi-stage ESP. The impactor data were used to calculate the penetration as a function of particle diameter for the multi-stage ESP.

The results of the experiments are shown in Figure 7. The data in Figure 7 were used to estimate the performance of an ESP that was converted to E-SOX. The calculations were performed by converting the penetration as a function of particle diameter data in Figure 7 to migration velocity as a function of particle diameter, see Figure 8. These data were then used to calculate the performance of an E-SOX ESP collecting fly ash with a typical size distribution (mass mean diameter 15 μm, geometric standard deviation 3.5) and E-SOX solids with a size distribution typical of those generated in our experiments (mass mean diameter 3 μm, and geometric standard deviation 3.5). The overall mass penetration of the ESP was determined by integrating over the two size distributions. The results

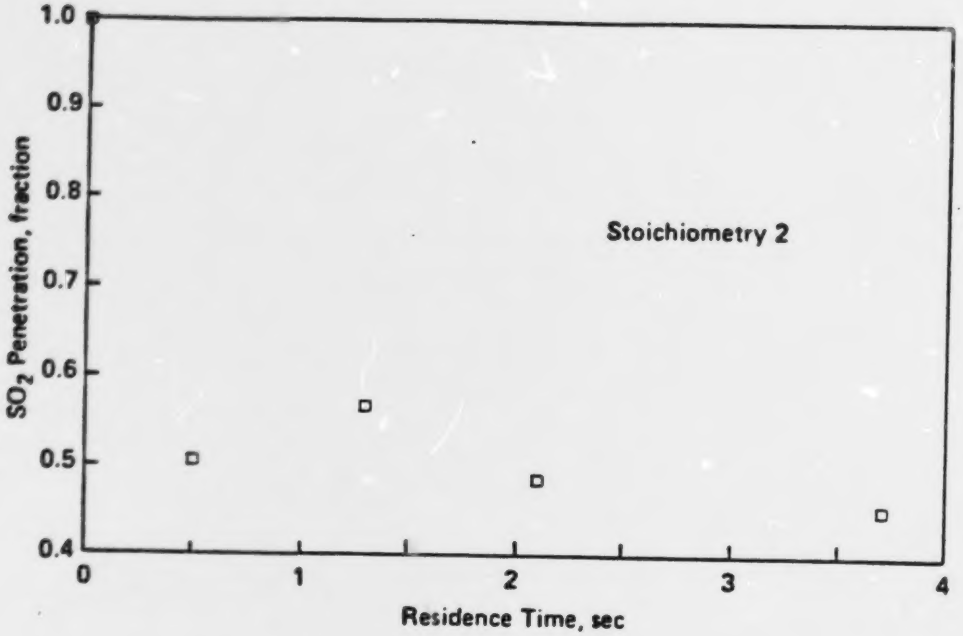


Figure 5. Penetration vs. time for lime.

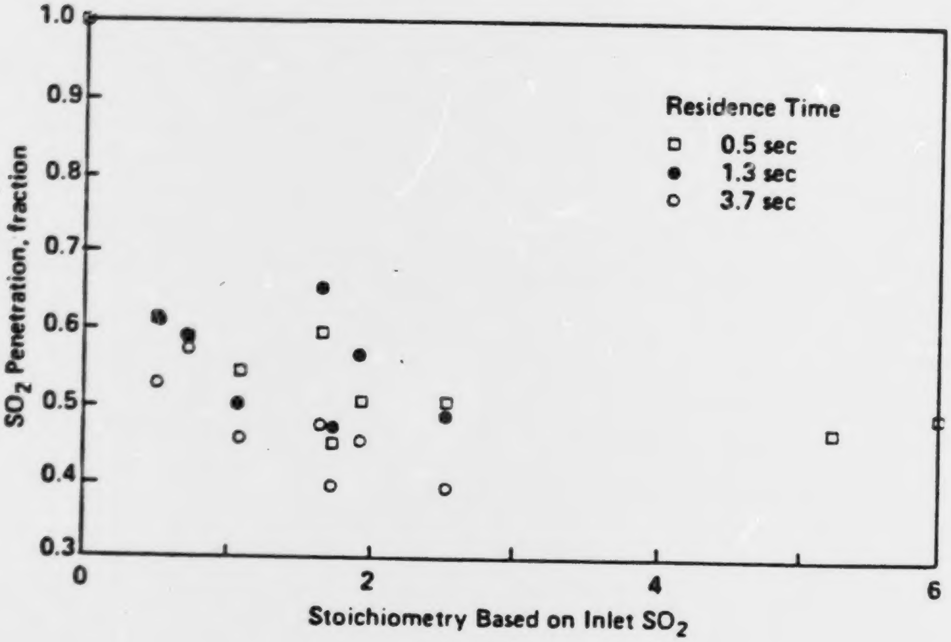


Figure 6. E-SOX data for lime.

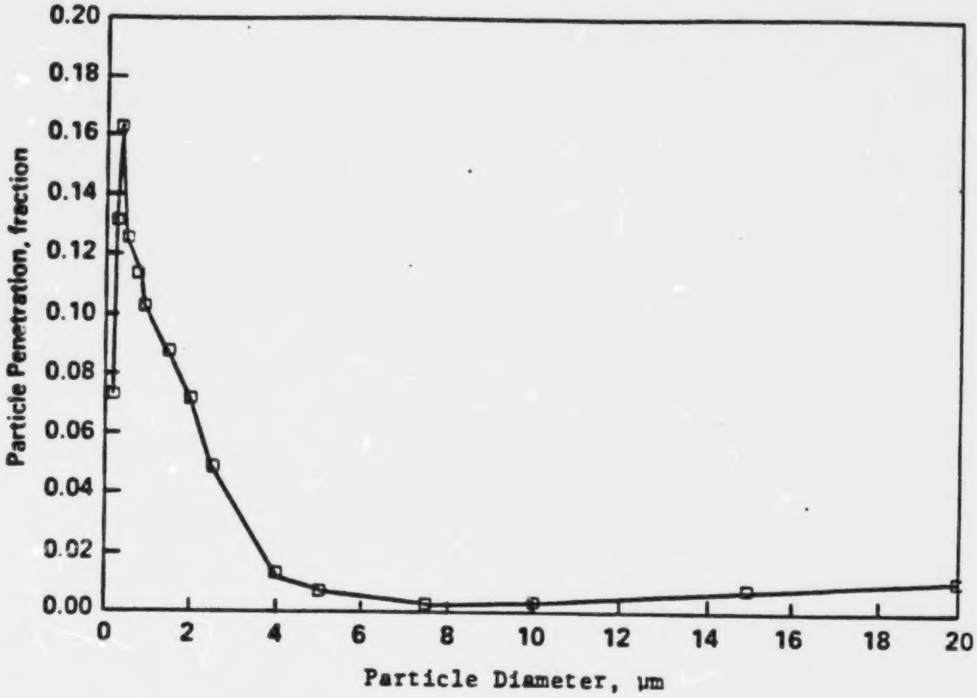


Figure 7. Results of E-SOX ESP experiments

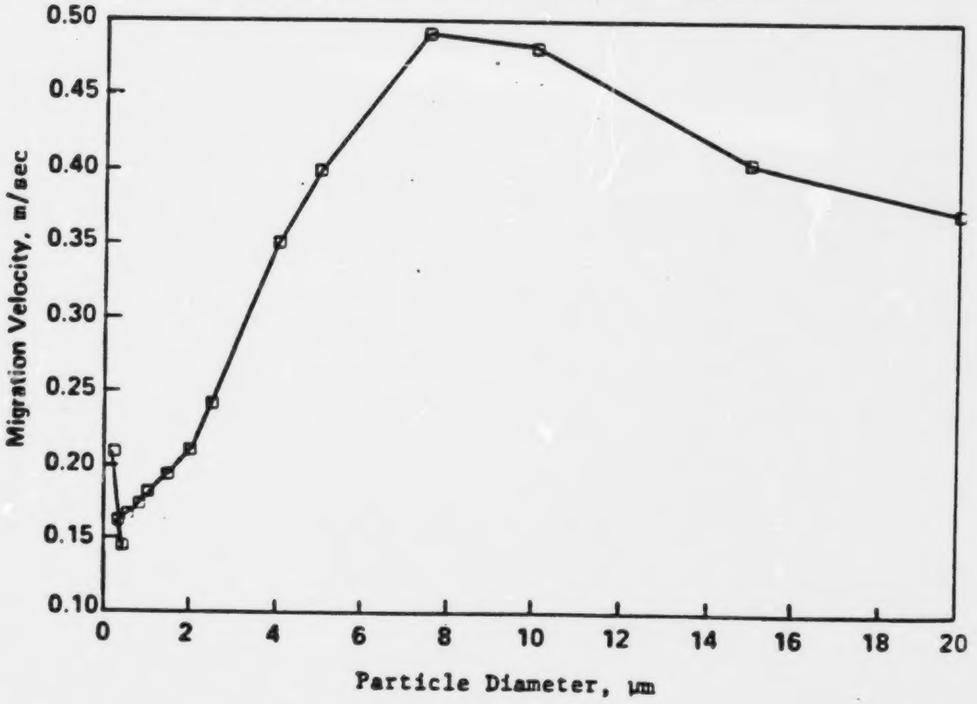


Figure 8. Migration velocity vs. particle diameter.

were corrected for sneakage and reentrainment using procedures similar to those used in the EPA/SoRI ESP model. (2)

The results of the particle collection calculations are shown in Figure 9. Note that the mechanical conditions of the ESP, as characterized by the uniformity of the gas flow distribution and the gas sneakage, are important factors. The mechanical conditions must be representative of current good practice if the particulate emission limitations are to be met.

ECONOMICS

A study estimate of the economics of E-SOX was conducted to determine the economic feasibility of the system. The capital costs of E-SOX were estimated by first analyzing the process diagram for the E-SOX system, Figure 10, to identify equipment that was required by other SO_x removal processes. This analysis showed that the lime handling system and the reacted lime waste disposal systems were common to most FGD processes. Because these two items, the lime handling system and the extra waste disposal system, are common to E-SOX and spray drying, the portion of the E-SOX capital cost due to these items was based on costs of the same two items for spray drying.

The capital costs of the cold-pipe precharger were based on experience gained from fabrication of a 5 MW pilot plant. The costs of the large diameter wire retrofit were based on discussions with nozzle vendors.

The cost analysis showed that the major costs were associated with the lime handling system followed by the costs of waste disposal--the two components with the firmest costs. The costs of the cold-pipe precharger, large diameter wires, and spray system were a small part of the total costs. Thus, even a factor of two error in estimating these costs would have little impact on the total economics of E-SOX.

The results of the capital cost study are shown in Figure 11. Also shown in Figure 11 are the costs of wet limestone FGD removal processes. Note that E-SOX is at least competitive with current technology and appears to offer significant cost advantages.

CONCLUSION

The experimental data and the modeling based on that data show that E-SOX is capable of providing 50-60% SO_2 removal in an existing ESP while meeting a particulate emission standard of 43 ng/J. The economic analysis shows that E-SOX has cost advantages when compared with other methods of SO_2 control.

Research is continuing to improve and verify the SO_2 and particulate removal models for E-SOX. Also a series of in-house experiments with the entire system of SO_2 removal and multi-stage ESP is planned for early 1985. A large pilot scale evaluation of E-SOX may take place in 1986.

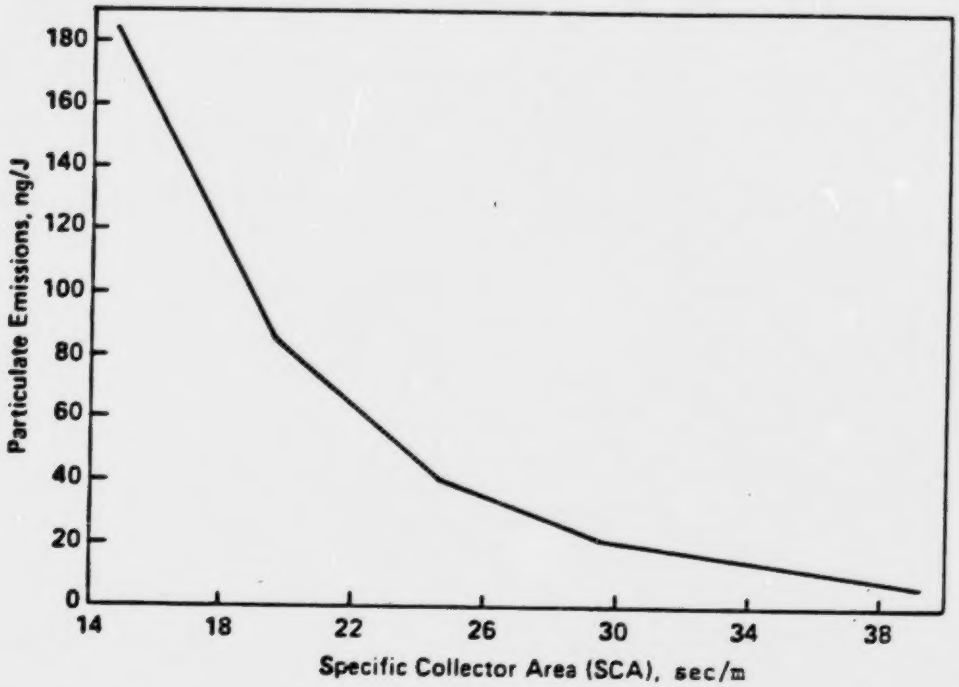


Figure 9. E-SOX predicted particulate emissions.

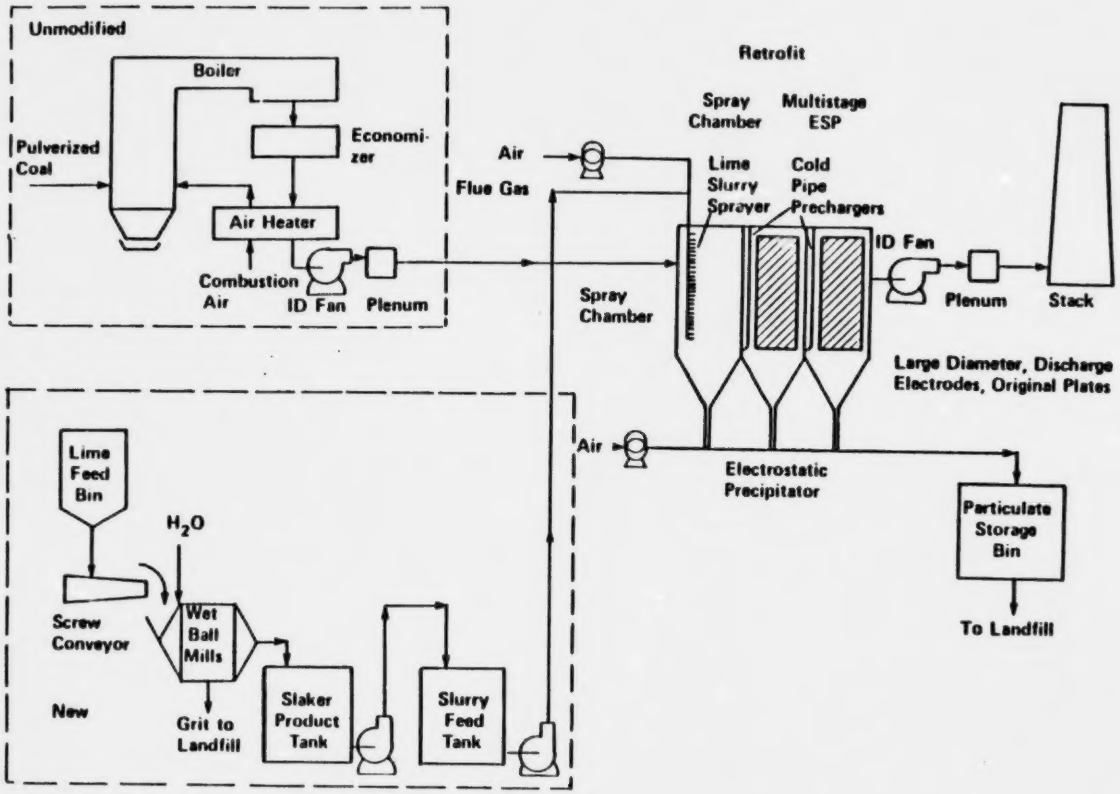

Unmodified
Boiler
Pulverized Coal
Economi-zer
Air Heater
Combustion Air
ID Fan
Plenum
Air
Flue Gas
Retrofit
Spray Chamber
Multistage ESP
Lime Slurry Sprayer
Cold Pipe Prechargers
ID Fan
Plenum
Stack
Spray Chamber
Large Diameter, Discharge Electrodes, Original Plates
Air
Electrostatic Precipitator
Particulate Storage Bin
To Landfill
Lime Feed Bin
H_2O
Screw Conveyor
Wet Ball Mills
Grit to Landfill
Slaker Product Tank
Slurry Feed Tank
New

Figure 10. Process flow diagram for E-SOX

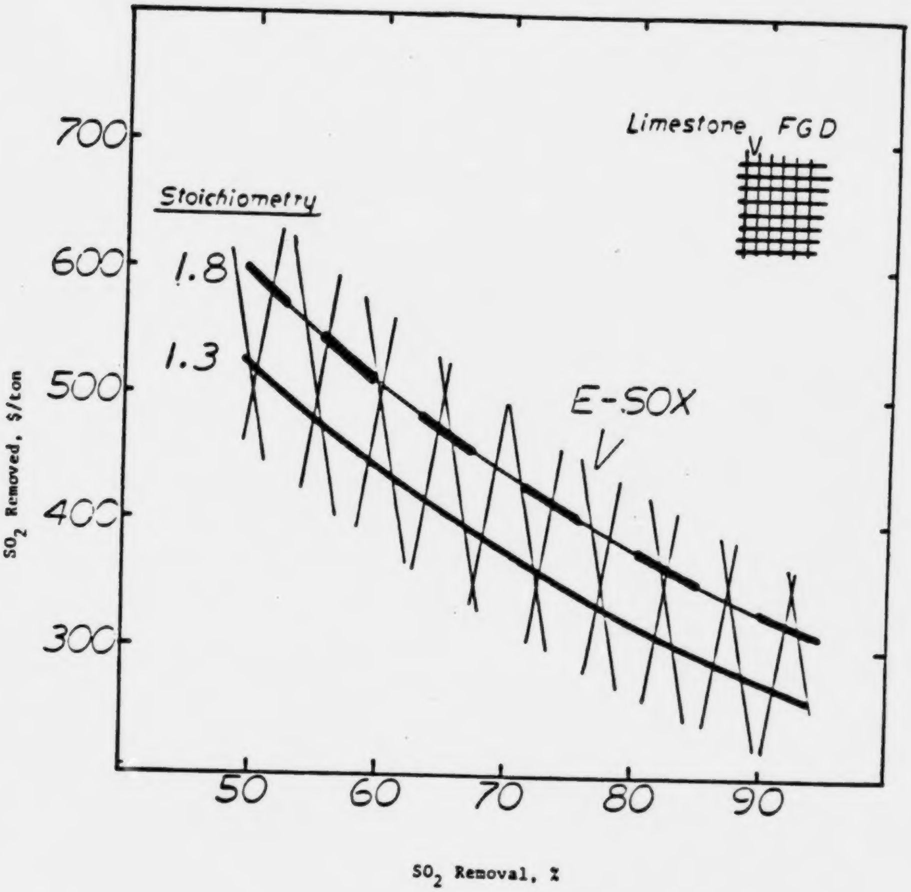


Figure 11. Cost effectiveness of E-SOX process

Research is also planned to reduce the amount of lime required for a given level of SO_2 removal. Reducing the amount of lime reduces the particulate burden on the multi-stage ESP and can significantly reduce the capital and operating costs of the system.

ACKNOWLEDGEMENTS

The economic analysis for E-SOX was conducted by John Milliken of AEERL.

REFERENCES

1. Damle, A.S. and L. E. Sparks, "Modeling of SO_2 Removal in Spray-Dryer Flue-Gas Desulfurization System." Paper presented at the Fifth Symposium on the Transfer and Utilization of Particulate Control Technology. U. S. EPA and EPRI, Kansas City, MO, August 27-30, 1984.

2. Faulkner, M. G. and J. L. DuBard, "A Mathematical Model of Electrostatic Precipitation (Revision 3)," Volumes I and II, and source code tape EPA-600/7-84-069a,b,c (NTIS PB84-212-679, PB84-212-687, PB84-232-990), U. S. EPA, IERL-RTP, N.C. 1984.

MITSUI-BF SIMULTANEOUS SOx AND NOx REMOVAL SYSTEM

by: Yoshiro Ito , T. Fujimoto and O. Nagaoka
Mitsui Mining Company, Limited
2-1-2, Nihonbashi-Muromachi, Chuo-ku, Tokyo, Japan 103

ABSTRACT

Mitsui Mining Company, Limited(MMC) has been developing the titled system by using activated coke jointly with Mitsui Miike Engineering Corporation(MMEC) by improving a process originated by Bergbau Forschung GmbH(BF) of West Germany. A pilot plant with a capacity of 1000 m^3 N/H has been operated from February, 1981 till September, 1983 at Tochigi Works of MMEC.

MMC has achieved success in the long term continuous operation of the pilot plant at an efficiency of more than 99% SOx removal and more than 80% NOx removal by the further improvement of BF System.

With the results of test performance, MMC has gained confidence in applying the new technology in a commercial plant. MMC constructed the titled system plant of 30,000 m^3 N/H capacity of MMC Coal-fired Power Station at Omuta, Japan. Its commercial operation began on October 1, 1984 and the plant has been operated satisfactorily so far.

MMC has been operating an activated coke (AC) production pilot plant of a capacity of 0.8 ton AC/day in an attemp to produce less expensive AC with better performance. This AC is used in the titled system plant at Omuta and has shown very satisfactory results. After performance test operation based on a design condition at 99% DeSOx and 50% DeNOx was achieved, higher NOx removal efficency operation(70%-80%) is started since February, 1985.

MITSUI-BF SIMULTANEOUS SOx AND NOx REMOVAL SYSTEM

GENERAL

Nowadays, simultaneous removal of sulfur oxides (SOx), Nitrogen oxides (NOx) and dust is required in the air pollution control systems. For the purpose of satisfying this requirement, Mitsui Mining Company (MMC) has been developing simultaneous SOx & NOx removal process, using activated coke jointly with Mitsui Miike Engineering Corporation (MMEC) by improving a process originated by Bergbau Forschung (BF) of West Germany. A pilot plant with a capacity of treating 1,000 Nm^3/H has been operated from February, 1981 till September, 1983 at Tochigi Works of MMEC.

MMC has achieved success in the long term continuous operation of the pilot plant at an efficiency of more than 99% SOx removal and more than 80% NOx removal by the further improvement of BF system.

With the results of test performance, MMC has gained confidence in applying the new technology in a commercial plant and had established Mitsui-BF Simultaneous SOx & NOx Removal System. MMC constructed a dry type SOx & NOx removal plant of 30,000 Nm^3/H capacity at MMC Coal-fired Power Station at Omuta City, Kyushu. Its commercial operation begun on October 1, 1984 has continued satisfactorily so far. This plant was constructed by MMEC as main contractor, and its operation was approved by MITI after official inspection on September 28, 1984, as the first plant of dry type simultaneous SOx & NOx removal plant.

Having developed the own process of producing the activated coke, MMC is operating an activated coke production plant having a capacity of 0.8 t/day in an attempt to produce less costly activated coke with sufficient activity and strength. This activated coke is being used at the dry type SOx & NOx removal plant at Omuta power station and is showing very satisfactory result.

PROCESS DESCRIPTION

Mitsui-BF Simultaneous SOx and NOx Removal System is shown in Figure 1.

SOx & NOx REMOVAL PROCESS

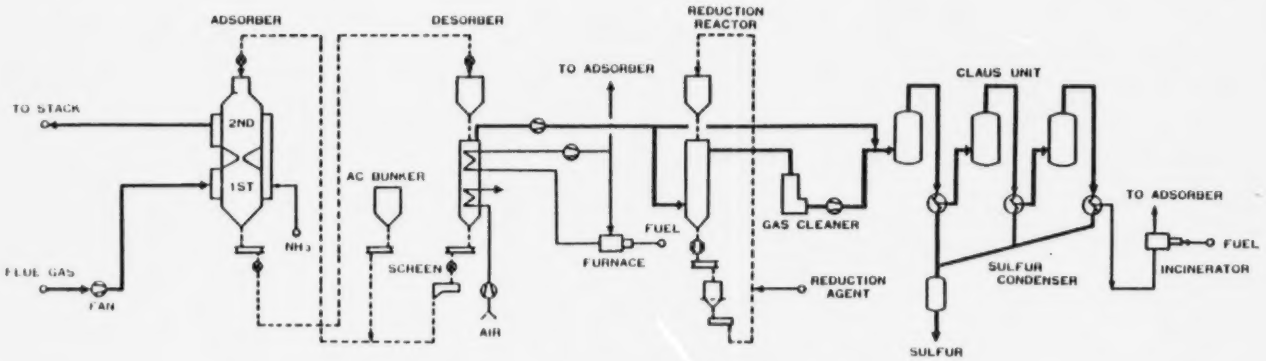

ADSORBER
DESORBER
REDUCTION REACTOR
TO ADSORBER
CLAUS UNIT
TO STACK
2ND
1ST
AC BUNKER
NH₃
FLUE GAS
FAN
SCREEN
AIR
FURNACE
FUEL
GAS CLEANER
REDUCTION AGENT
TO ADSORBER
FUEL
INCINERATOR
SULFUR CONDENSER
SULFUR

Figure 1. **Mitsui-BF Simultaneous SOx and NOx Removal System**

Adsorption Section

Flue gas enters moving bed adsorbing system which holds activated coke. Normally flue gas temperature from air-heater ranges 120°c to 160°c. And this process produces the best efficiency in this temperature range. The adsorption system consisits of two stage vertical adsorbers where activated coke flows down by gravity from the top of the 2nd stage adsorber to the bottom of the 1st stage adsorber. On the other hand, flue gas horizontally passes through the 1st stage adsorber and goes into the 2nd stage adsorber. Mainly SOx removal is effected in the 1st stage adsorber and NOx removal, by injection of ammonia, is effected in the 2nd stage adsorber. Chemical reactions in the adsorption section are as follows:

SO_2 removal: $2SO_2 + O_2 + 2H_2O \longrightarrow 2H_2SO_4$ (1)

NOx removal: $4NO + O_2 + 4NH_3 \longrightarrow 4N_2 + 6H_2O$ (2)

Sub-reaction:

$NH_3 + H_2SO_4 \longrightarrow NH_4HSO_4$ (3)

$2NH_3 + H_2SO_4 \longrightarrow (NH_4)_2SO_4$ (4)

Sub-reactions (3) and (4) are dominant and so ammonia injection to flue gas containing high SOx does not bring high NOx removal efficiency. In other words, it is preferable to cause NOx removal reaction under the condition in which flue gas contacts small amount of SOx and proper amount of ammonia contacts activated coke. For this reason Mitsui-BF process uses two-stage adsorber.

Dust load in outlet flue gas of this adsorber is less than 20 mg/Nm^3 and it is possible to send the flue gas to stack without any dust collectors.

Desorption Section

Activated coke is discharged from the 1st stage adsorber and sent to the desorber for regeneration. Activated coke is regenerated by indirect heating at the temperature of 400°c ± 20°c. As a result of regeneration, about 33 vol. percent of enriched SO_2 gas is generated from the desorber. Chemical reactions in the desorption are as follows:

$H_2SO_4 \longrightarrow H_2O + SO_3$ (5)

$2SO_3 + C \longrightarrow 2SO_2 + CO_2$ (6)

$3SO_3 + 2NH_3 \longrightarrow N_2 + 3SO_2 + 3H_2O$ (7)

$(NH_4)_2SO_4 \longrightarrow 2NH_3 + SO_3 + H_2O$ (8)

$$2NH_3 + 3SO_3 \longrightarrow N_2 + 3SO_2 + 3H_2O \qquad (9)$$

The regenerated activated coke is indirectly cooled by air and recycled to the adsorber.

Sulfur Recovery Section

This section consists of a unit to convert SO_2 generated in the desorber to H_2S and Claus unit. Two thirds of SO_2 enriched gas is introduced to the coke reactor where SO_2 is reducted to H_2S with hot carbonaceous material and steam at the temperature of 650°c to 950°c. H_2S from the coke reactor is mixed with the residual enriched SO_2 gas and sent to Claus unit. Representative reactions in the coke reactor can be expressed as follows:

$$C + SO_2 \longrightarrow S + CO_2 \qquad (10)$$

$$C + CO_2 \longrightarrow 2CO \qquad (11)$$

$$C + H_2O \longrightarrow CO + H_2 \qquad (12)$$

$$2H_2 + S \longrightarrow 2H_2S \qquad (13)$$

$$2CO + S \longrightarrow 2COS \qquad (14)$$

Mixture of reduced gas and by-pass gas becomes elemental sulfur in the Claus unit in accordance with the Claus reaction formula (15).

$$2H_2S + SO_2 \longrightarrow 3S + 2H_2O \qquad (15)$$

COS is also recovered finally as elemental sulfur as by the equation (16).

$$COS + H_2O \longrightarrow H_2S + CO_2 \qquad (16)$$

The tail gas from the Claus unit is introduced to an incinerator to convert residual elemental sulfur and H_2S to SO_2 and fed to the 1st stage adsorber.

FEATURES OF THE PROCESS

Mitsui-BF simultaneous SOx & NOx removal system has the following features:

(1) SOx removal efficiency as high as 98% or more.

(2) NOx removal efficiency at more than 80%.

(3) Long term continuous and stable operation of the system.

(4) Low cost of construction and operation.

(5) Small site of construction.

(6) No requirement of process water and waste water treatment system.

(7) No need for flue gas reheating.

(8) Removal of SO_3 contained in the flue gas as well as SO_2.

(9) High purity elemental sulfur as by-product (99.9% or more in form of molten or pellet).

(10) Alternative by-product in form of sulfuric acid or liquefied SO_2.

OPERATION OF TOCHIGI PILOT PLANT

SOx & NOx REMOVAL PROCESS

During the operation from February 1981 till September 1983, the pilot plant kept its stable running, and showed quite a satisfactory performance, as follows:

1.	Volume of flue gas handled	1000 Nm^3/H
2.	Temperature of flue gas	130°c
3.	Inlet concentration of SO_2	1000 ppm
4.	Inlet concentration of NOx	180 ppm
5.	SO_2 removal efficiency	99%
6.	SO_3 removal efficiency	98%
7.	NOx removal efficiency	85%
8.	Leak ammonia	15 - 20 ppm
9.	Feed NH_3	260 ppm
10.	Quantity of dust at outlet	20 mg/Nm^3 (5mg/Nm^3 with Bag Filter)
11.	Pressure drop across adsorption system	150 mm H_2O

ELEMENTAL SULFUR RECOVERING PROCESS

This process was developed by Mitsui Mining Company and Mitsui Miike

Engineering Corporation by their own efforts. From its first operation in April 1982, elemental sulfur with high purity was produced at high sulfur yield under the following conditions:

1.	Volume of gas handled	30 Nm^3/H
2.	Inlet concentration of SO_2	20 Vol.%
3.	Reducing agent	Coke
4.	Reducing temperature	650°c - 950°c
5.	Purity of recovered sulfur	99.9% or more

COMMERCIAL OPERATION OF SOx & NOx REMOVAL PLANT AT MMC COAL-FIRED POWER STATION IN OMUTA

August 23, 1984 - Boiler flue gas was introduced to adsorber and SOx removal operation was started.

September 5 - Injection of ammonia was begun to start the simultaneous SOx & NOx removal operation.

September 9 - Enriched SO_2 gas was introduced into by-production system to start the total system operation.

September 28 - Governmental inspection (MITI) and permit was obtained.

October 1 - Commercial operation started.

Since then, the dry type SOx & NOx removal plant including by-production system is continuously running satisfactorily. Higher NOx removal operation is started in this year.

Actual operating data of the performance test run of Omuta plant are summarized in Table 1.

Actual operating data of the higher NOx removal efficiency operation of Omuta plant are summarized in Table 2.

Actual Utility Consumption in case of performance test run of Omuta plant is summarized in Table 3.

TABLE 1. ACTUAL OPERATING DATA OF THE PERFORMANCE TEST RUN OF OMUTA PLANT

OPERATING PERIOD: FROM OCT.1ST TILL NOV.20, 1984

		Actual data	Design data
1.	Inlet gas condition		
	Flue gas volume	$30,000m^3$ N/H	$30,000m^3$ N/H
	Temperature	120 - 140°c	120°c
	SO_2 concentration	480 - 550ppmD.B.	550ppmD.B.
	NOx concentration	140 - 165ppmD.B.	172ppmD.B.
2.	Outlet gas condition		
	SO_2 concentration	trace	trace
	NOx concentration	70 - 87ppmD.B.	94ppmD.B.
	Leak ammonia	5 - 10ppmD.B.	10ppmD.B.
	Dust load	15 - $20mg/m^3$ N	$20mg/m^3$ N
3.	Activated coke	Mitsui AC	Mitsui AC
4.	Loss of AC	1.8 - 2.0% of AC flow rate	2.2% of AC flow rate
5.	Pressure drop of adsorber	100mmAqG	100mmAqG
6.	SO_2 enriched gas		
	Flow rate	55 - $65m^3$ N/H	$63.7m^3$ N/H
	SO_2 concentration	22 - 23 vol.%	22.4vol.%
	H_2O concentration	50 - 53	53 vol.%
	N_2, CO_2 concentration	balance	balance
7.	Convertion rate		
	sulfur recovery section	90 - 93%	89%
8.	By-product sulfur		
	Quantity	460kg/day	460kg/day
	Purity	99.9% or more	99.9% or more

TABLE 2. ACTUAL OPERATING DATA OF THE HIGHER NOx REMOVAL EFFICIENCY OPERATION OF OMUTA PLANT

OPERATING PERIOD: FROM FEB.6, 1985 -	
Inlet gas condition	
Flue gas volume	20,000 m^3 N/H
Temperature	140 - 150°c
SO_2 concentration	700 - 800 ppm
NOx concentration	145 - 170 ppm
Outlet gas condition	
SO_2 concentration	trace
NOx concentration	29 - 40ppm D.B.
Leak ammonia	10 - 25ppm D.B.
Dust load	15 - 20 mg/m^3N
DeNOx efficiency	75 - 80%
Activated coke	Mitsui AC

TABLE 3. ACTUAL UTILITY CONSUMPTION IN CASE OF PERFORMACE TEST RUN OF OMUTA PLANT

	Actual data	Design data
Electric power	142kwH/H	150kwH/H
Activated coke	8 - 9kg/H	10kg/H
Heavy oil	18 ℓ /H	20 ℓ /H
Ammonia	2.1kg/H	2.9kg/H
Nitrogen #	10kg/H	9kg/H

# Nitrogen is used for sealing SO_2 leakage from rotary value of desorption section to adsorption section.

POTENTIAL APPLICATION OF THE DRY TYPE SOx & NOx REMOVAL SYSTEM TO OTHER THAN POWER STATIONS TO THE FOLLOWING INDUSTRIES

EXHAUST GAS OF FCC(FLUID CATALYTIC CRACKING) PLANT OF PETROLEUM REFINERY

Volume of exhaust gas	200,000 - 300,000 Nm^3/H
Temperature of exhaust gas	180°c - 260°c
SOx	200 - 700 ppm
SO_3 in SOx	14 - 50%
NOx	200 - 400 ppm
Dust load	60 - 200 mg/Nm^3
SOx removal efficiency	80 - 90% or more
NOx removal efficiency	65 - 80% or more
Outlet dust load	30 - 50 mg/Nm^3 or less

High SO_3 removal efficiency of this process will bring about excellent performance for this application.

EXHAUST GAS OF WASTE LIQUOR INCINERATOR

Volume of exhaust gas	4000 Nm^3/H
Temperature of exhaust gas	230°c - 240°c

Composition of exhaust gas:

CO_2	10.7 vol.%
O_2	4.5 vol.%
H_2O	10.9 vol.%
N_2	73.9 vol.%
NOx	820 - 1640 ppm
SOx	less than 5 ppm

Dust:

NaCl	20 mg/Nm^3
NaF	10 mg/Nm^3
Na_2SO_4	10 mg/Nm^3
Outlet Nox	150 ppm or less
Outlet dust load	30 mg/Nm^3

Because of the poisoning effect to catalyst by NaCl, NaF and Na_2SO_4 contained in the dust, conventional SCR system is not so feasible without an extra dust removal system.

EXHAUST GAS OF VARIOUS PETROCHEMICAL PLANTS

In recent Japan, exhaust gas of petrochemical plants is frequently being fed to gas turbine to recover the enthalpy. In this process, cleaning of exhaust gas at the inlet of gas turbine by use of SOx & NOx removal plant is being demanded. Following is an example of data of exhaust gas.

Gas volume	60,000 Nm^3/H
Gas temperature	210°c - 220°c
SO_2	890 ppm
SO_3	180 ppm
NOx	360 ppm
NH_3	100 - 127 ppm
CO	7.9 vol.%

Even in this case, conventional wet scrubbing system is not feasible because it does not effectively remove SO_3 and causes temperature drop of the gas.

EXHAUST GAS TREATMENT AT SINTERING PLANT OF STEEL WORKS

Typical composition of exhaust gas at sintering plants is described as follows:

SO_2	150 - 250 ppm
NOx	150 - 200 ppm

O_2	10 - 15%
H_2O	10 - 12%
N_2	Balance

Process to efficiently remove SOx and NOx from the above exhaust gas is now being studied by us.

NOTE

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

OPERATING EXPERIENCES AND DEVELOPMENTS WITH LJUNGSTROEM GAS-GAS HEATERS IN WEST GERMAN FGD PLANTS

by: David E. Clay
Kraftanlagen Aktiengesellschaft Heidelberg
Heidelberg, Federal Republic of Germany.

ABSTRACT

Regenerative gas-gas heaters have been adopted by the German Power Industry as the state of the art for reheating scrubbed flue gases.

The anticipated fouling and corrosion problems have proved controllable and the first large-scale gas-gas heater entered operation in 1982 at NWK Wilhelmshaven. Since then more than 11.000 operating hours have been achieved.

Two novel features are incorporated in the design - a cold side plastic heating element and an on-line high-pressure washing system.

Anticipated future SO_2 emission limits of 200 mg/Nm^3 have led to a low leakage design (0.3-0.5 %).

The operating experiences and developments with the regenerative gas-gas heater will be discussed.

INTRODUCTION

In 1977 the first German large-scale prototype flue gas desulfurisation plant, based on a wet process, was commissioned at the 720 MW power station owned by Nordwestdeutsche Kraftwerke Aktiengesellschaft in Wilhelmshaven, West Germany.

During the period 1977 to 1980, pilot trials were carried out on reheating systems for the treated gas. Reheating has a considerable influence on both capital investment costs and, perhaps more importantly, on plant operating costs.

At Wilhelmshaven various reheating systems were tested, e.g. steam coils, heat pipes, recuperative and regenerative heat exchangers.

After 4 years of pilot trials the management at NWK decided to install a Ljungstroem regenerative gas-gas heater in their first full-scale FGD plant (untreated gas flow 1.3 x 10^6 Nm^3/h).

The scrubber and gas-gas heater entered service on March 1st, 1982, and since then more than 11.000 operating hours have been achieved. At no time during this period has the gas-gas heater caused a plant shut-down or the plant to operate at reduced load.

PILOT TRIAL RESULTS

The results of the pilot trials have been previously reported (1). As a summary it can be said that from the large number of materials tested it was decided to select Corten as the structural material, a thermoplastic material for the cold side heating elements and an enameled steel element at the hot side. Corrosion rates were measured during the trials and it was decided to incorporate a corrosion allowance into the rotor structure and line stationary components with proprietary flake glass or rubber linings.

THE WILHELMSHAVEN GAS-GAS HEATER DESIGN

The Wilhelmshaven gas-gas heater was designed to reheat 1.5 x 10^6 Nm^3/h of treated gas from 54°C to 100°C (Table 1), and the rotor diameter is approx. 16 m.

TABLE 1. GAS-GAS HEATER DESIGN DATA FOR NWK WILHELMSHAVEN

Untreated gas flow (Nm^3/h)	:	1.300.000
Inlet temperature (°C)	:	135
Outlet temperature (°C)	:	81
Design pressure drop (mmWG)	:	51
Treated gas flow (Nm^3/h)	:	1.500.000
Inlet temperature (°C)	:	54
Outlet temperature (°C)	:	100
Design pressure drop (mmWG)	:	55
Untreated gas leakage (Nm^3/h)	:	38.000
Heating element height		
Hot side - enameled (mm)	:	900
Cold side - plastic (mm)	:	300
Minimum element temperature (°C)	:	57
Total heating transfer surface (m^2)	:	41.750

The gas-gas heater, Figure 1, comprises the same basic components as a normal air preheater, namely

- structural housing
- rotor
- heating elements
- sealing system
- drive units
- cleaning device.

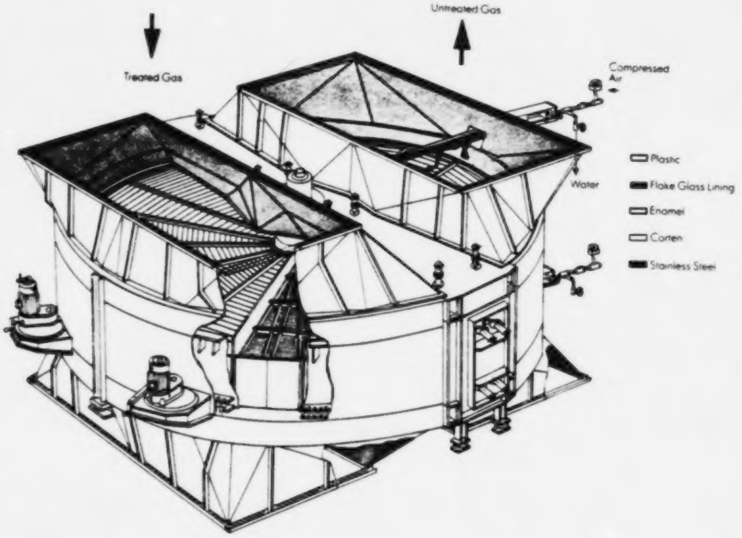

Untreated Gas
Treated Gas
Compressed Air
Water
▭ Plastic
▬ Flake Glass Lining
▭ Enamel
▭ Corten
▬ Stainless Steel

Figure 1. Ljungstroem Gas-Gas Heater.

OPERATING EXPERIENCES

The main areas of concern for regenerative gas-gas heaters are corrosion and fouling.

CORROSION

Corrosion attack is to be expected at both the untreated and treated gas sides of the gas-gas heater.

At the untreated gas side the gas is cooled from typically 130°C to 80°C, i.e. through the acid dew point.

At the treated gas side the residual SO_2/SO_3 in combination with the high water content of the treated gas and chlorides in the residual droplets also forms a corrosive environment.

After approximately 10.000 operating hours, measurements of corrosion rates at the radial walls of rotor have shown greatly reduced rates from those extrapolated from the pilot trials. This is considered to result from a vastly improved demister design and pre-drying of the treated gas leaving the scrubber.

No corrosion attack has been found at the plastic, although age hardening has occurred.

Slight corrosion has been found at the enameled elements, although this is not considered to make replacement necessary for another 3 years.

The rubber lining at the inlet treated gas duct and the glass flake lining at the treated gas outlet/untreated gas outlet ducts showed no signs of corrosion attack.

Test lining of rotor parts with a polyamide paint proved unsuitable and large areas had flaked away. However, no significant corrosion could be detected.

FOULING

Fouling occurs in the gas-gas heater due to:

- residual fly ash depositing onto the heating element surface at the untreated gas side
- evaporation of the residual droplets from the demister onto the cold side heating elements leaving the dissolved and undissolved solids behind as a deposit. This deposit only penetrates 15-20 mm into the matrix depth, Figure 2.

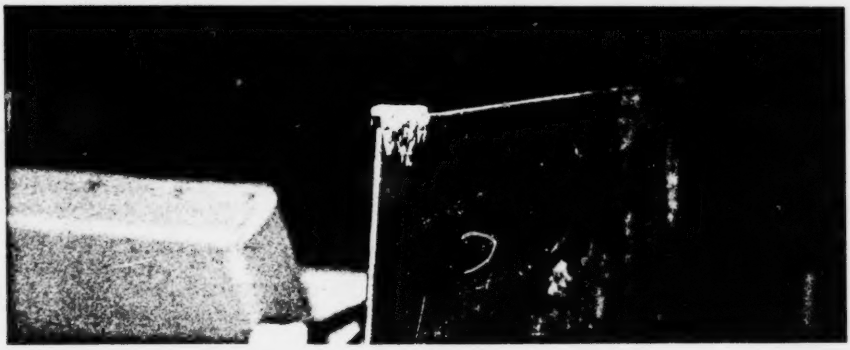


Figure 2. Fouling Profile.

The residual fly ash content can have a major effect on the rate of pressure drop rise.

At Wilhelmshaven tests were carried out to determine this effect by removing compartments from the electro-static precipitator until dust loadings of more than 200 mg/Nm^3 were obtained. At this dust loading the pressure drop rose rapidly in spite of sootblowing.

It was therefore decided to test the high-pressure water washing equipment over a small radial band of the rotor. The result was highly encouraging, Figure 3, and the decision was made to try a full-scale test.

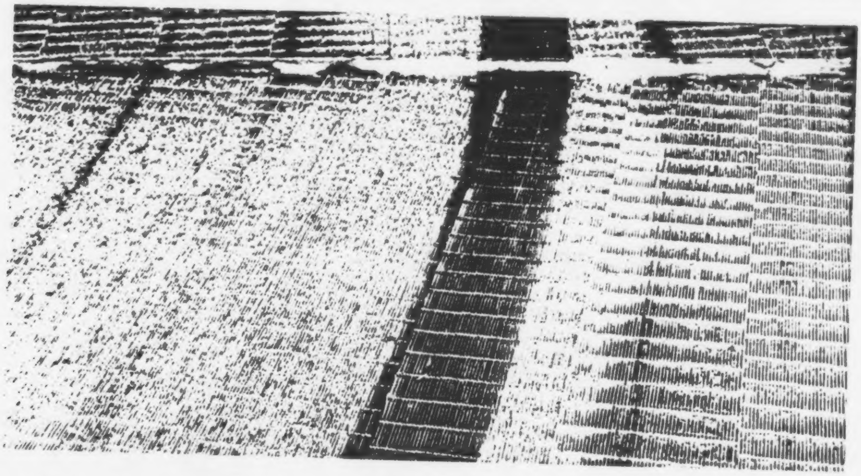


Figure 3. Result of HP Water Washing Test.

Prior to the test a number of enameled heating elements were removed in order to determine the degree of fouling, Figure 4.

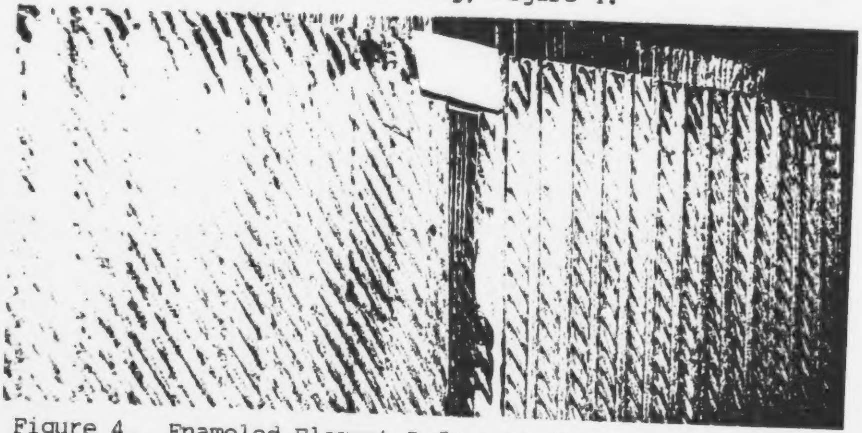


Figure 4. Enameled Element Before HP Washing.

Reproduced from
best available copy.

The full-scale test was carried out using water at 80 bar from the hot (enameled) side and 40 bar at the cold (plastic) side, Figure 5.

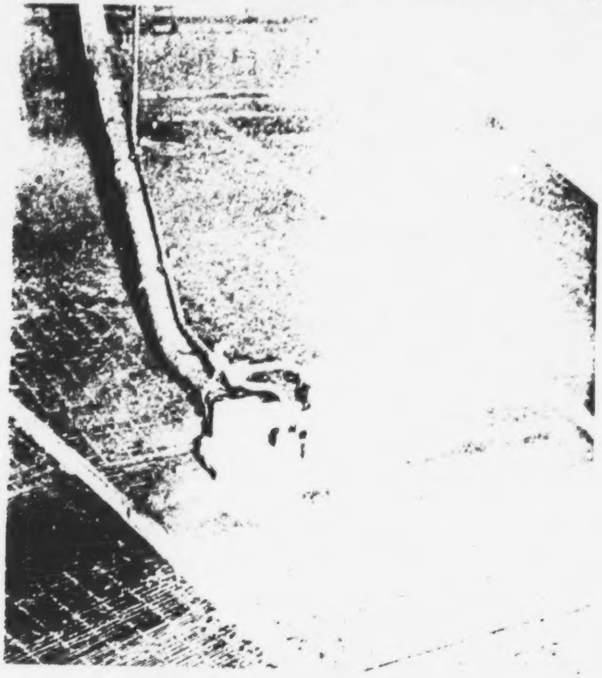


Figure 5. High-Pressure Water Washing.

The results were so successful, Figure 6, that "on-line" washing was adapted as a standard cleaning procedure when the pressure drop over the heat exchanger increases by 30 % over the design value.

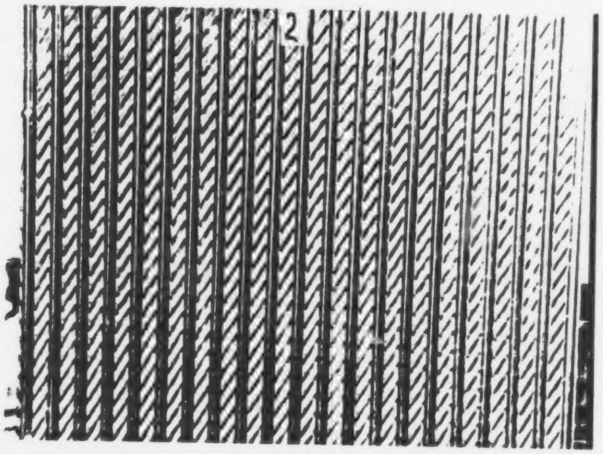


Figure 6. Enameled Element after HP Washing.

The residual water droplet content, the size of the droplets and the total salt content also play an important role in the rate of fouling.

With a residual droplet content of more than 100 mg/Nm^3 it is highly recommended that the treated gas be "pre-dried" after leaving the demister. At Wilhelmshaven this was carried out using hot, untreated gas, which led to a decrease in the overall plant De-SOx efficiency. As a result, the power station management gradually decreased the pre-drying rate from 8°C to zero pre-drying and has operated the gas-gas heater without pre-drying since 1983. This has resulted in an increased high-pressure water washing frequency, Figure 7. It is thus recommended that pre-drying be made using hot treated gas.

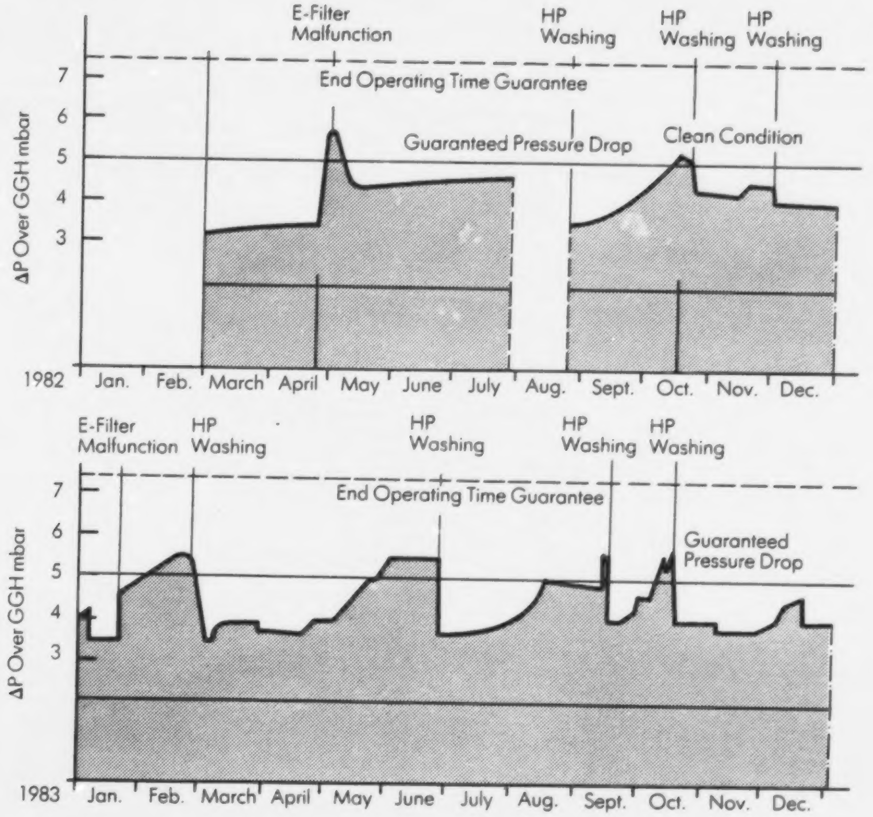


Figure 7. Pressure Drop Profile.

It is interesting to note that at another FGD plant in Germany, where the droplet content is less than 40 mg/Nm^3 and the residual fly ash is less than 40 mg/Nm^3 the first 2000 operating hours have been achieved without the need for high-pressure water washing, in fact the pressure drop has only increased by approx. 10 %.

FURTHER DEVELOPMENTS IN GAS-GAS HEATER DESIGN

A number of design improvements have been made based on the excellent operating experience at Wilhelmshaven.

ENAMEL

In conjunction with a German enamel company, a new enamel has been formulated which has a corrosion resistance more than 100 % higher than enamels used in conventional air preheaters.

PLASTIC ELEMENTS - Figure 8

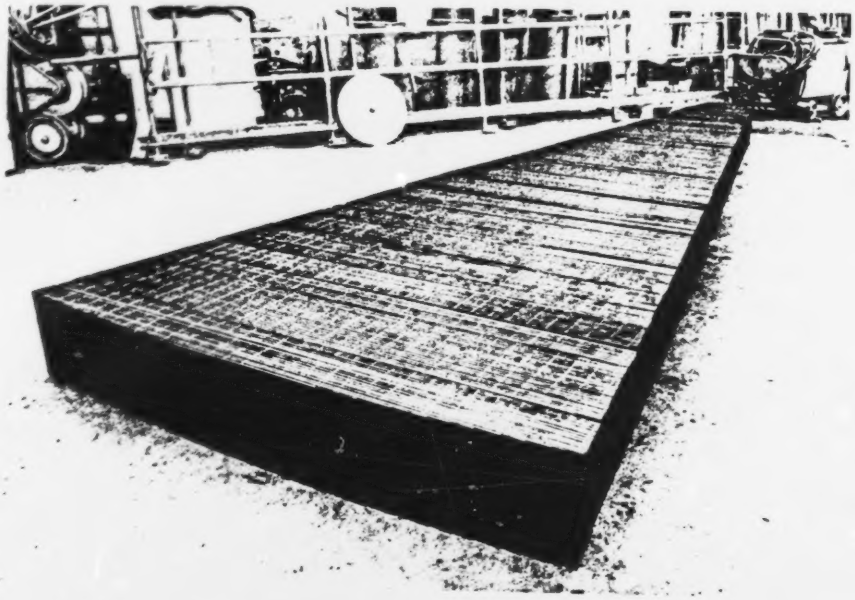


Figure 8. Plastic Heating Elements.

A reinforced plastic has been developed which has a higher continuous operating temperature combined with reduced age-hardening. Although plastics are available for 160-200°C temperatures, tests carried out at Wilhelmshaven have shown that these plastics become very brittle due to acid/temperature influences and manufacturing techniques.

COMBINED CLEANING DEVICE - Figure 9

An extra nozzle manifold has been added to reduce cleaning times and to increase the residence time.

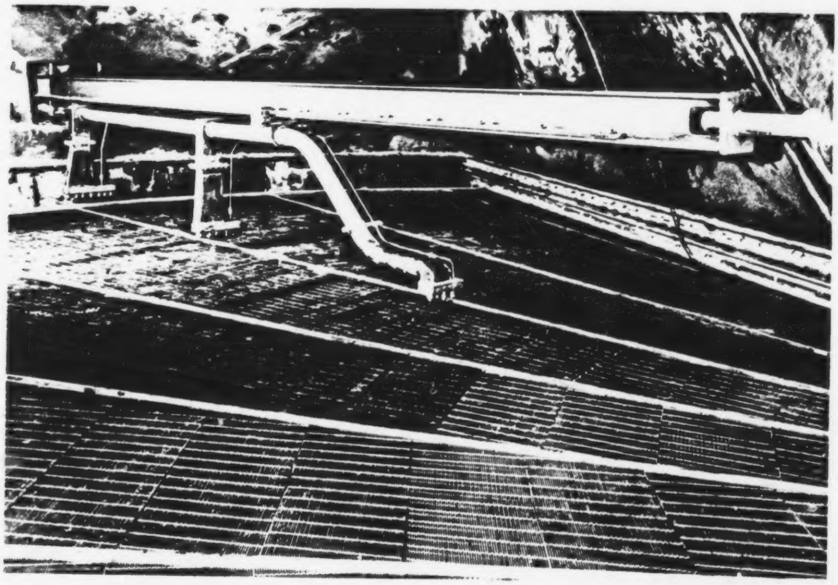


Figure 9. Combined Cleaning Device .

The retractable device is used for cleaning with compressed air and the same nozzles are used for low-pressure water washing. A separate nozzle group mounted adjacent to the manifolds is used for the HP water washing.

AXIAL SEAL

An axial seal has been added to the overall sealing system to reduce leakage rates further. The sealing system is then identical to that used by Kraftanlagen in large-sized air preheaters except that the radial seals are 1-piece units instead of 3- or 5-piece units as with air preheaters.

RADIAL SEAL CONTROL - Figure 10

Proximity switches are installed and used to control the gap at the hot side radial seal via electric actuators. This ensures rapid deflection of the seals to the thermal deformation of the rotor.

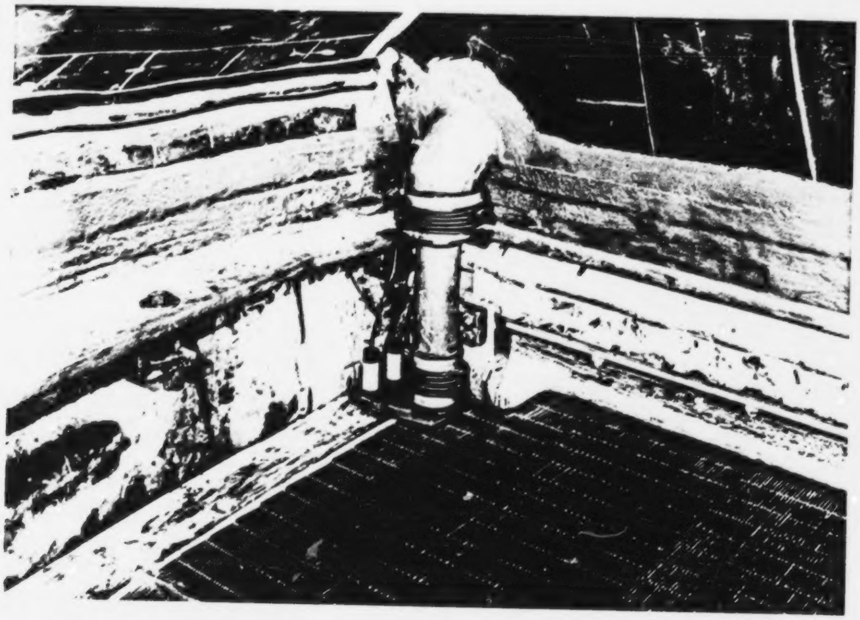


Figure 10. Proximity Switch Assembly.

LEAKAGE RATES IN REGENERATIVE GAS-GAS HEATERS

Regenerative gas-gas heaters have a system-inherent leakage which is made up by 2 components, namely:

Direct Leakage

under the radial and axial seals from the high-pressure side to the lower-pressure side.

Entrained Leakage

due to that volume of gas contained in a rotor sector which is carried through to each side of the gas-gas heater during each rotation.

The direction of the leakage will have a negative effect on the overall plant SO_2 removal efficiency if the flow is from the untreated gas side to the treated gas side.

Two solutions are available to minimise this effect:

- positioning of the booster fan
- the KAH leakage reduction system.

FAN ARRANGEMENT AND LEAKAGE DEPENDENCE - Figure 11

The position of the booster fan has a major influence on the total contamination of the treated gas by untreated gas.

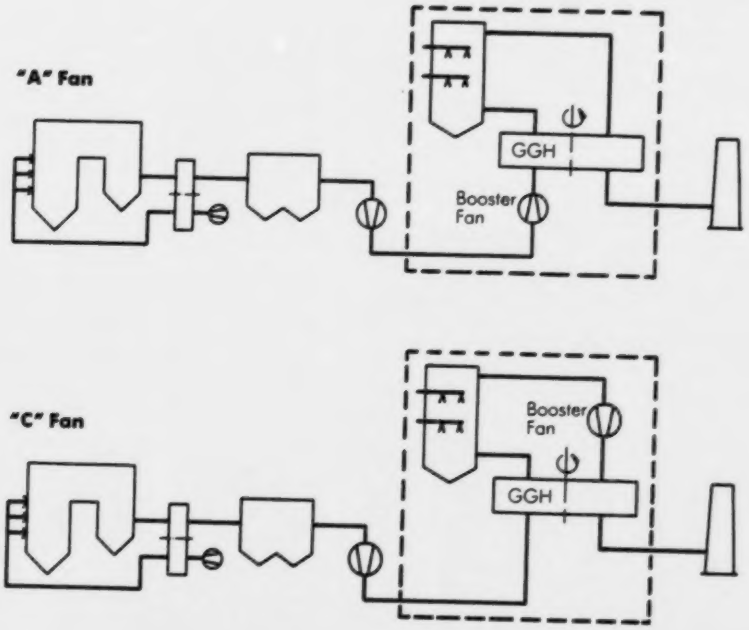


Figure 11. Booster Fan Arrangements.

'A' Fan Arrangement

The booster fan is installed in the hot untreated gas ducting. This arrangement results in leakage from the high-pressure untreated gas side to the lower-pressure treated gas side. The untreated gas leakage is typically 2-4 %.

The untreated gas leakage may only be reduced by the KAH leakage reduction system.

'C' Fan Arrangement

The booster fan is installed in the cold treated gas ducting. This arrangement results in leakage from the higher-pressure treated gas side to the untreated gas side.

Treated gas contamination is by entrained leakage only and is typically $\leqslant$ 1 %.

THE KAH LEAKAGE REDUCTION SYSTEM - Figure 12

The KAH leakage reduction system involves purging each rotor sector in the untreated gas side with hot treated gas immediately before rotation into the treated gas side, to reduce the entrained leakage and, indirectly, the direct leakage. Purging is made in the direction of untreated gas flow.

At the top side of the gas-gas heater a further flow of hot treated gas is blown through slits in the radial sector plate to form a pneumatic barrier to the untreated gas to further eliminate direct leakage.

Untreated gas leakage rates may be reduced to 0.5 - 0.8 % depending on the gas-gas heater size.

The KAH leakage reduction system may be designed for retrofitting, however, the 'passive' requirement must be incorporated in the original gas-gas heater design. This involves increasing the number of radial walls, widening and making slits in the radial seal plates, preparing weld-free supports for the purging skirt and installing take-offs into the top cover. The take-offs and slits are blanked off during passive operation.

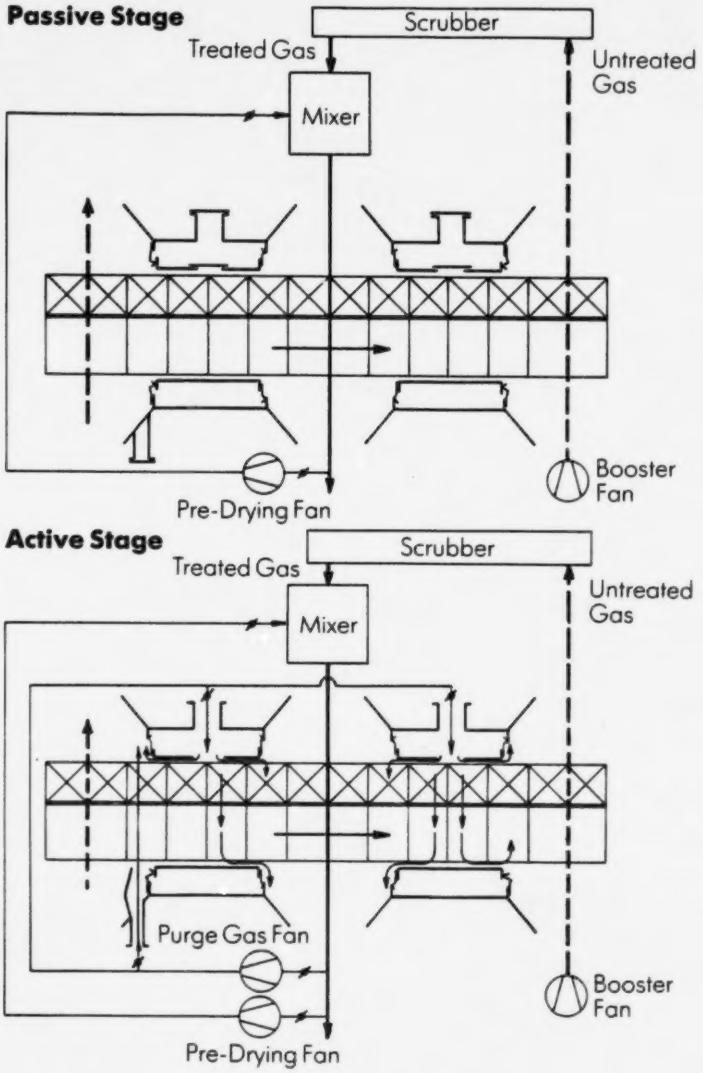

Passive Stage
Scrubber
Treated Gas
Untreated Gas
Mixer
Booster Fan
Pre-Drying Fan
Active Stage
Scrubber
Treated Gas
Untreated Gas
Mixer
Purge Gas Fan
Booster Fan
Pre-Drying Fan

Figure 12. KAH Leakage Reduction System .

Retro-fitting to 'active' operation involves modifications to the static seals of the radial sealing plates, the addition of a purging skirt, removing the blind flanges from the radial seals, installing the purge gas fan and connecting the intermediate ducting. Blind flanges should already be installed in the treated gas ducting. The tie-in work can be achieved in 3-5 days.

The purging system only may also be installed in gas-gas heaters for 'C' fan arrangements to reduce the entrained untreated gas leakage to $\leqslant 0.5$ %.

CONCLUSIONS

The regenerative gas-gas heater has proved to be a highly reliable component in flue gas desulfurisation plants.

The Kraftanlagen gas-gas heater design has shown that fouling and corrosion problems can be controlled without the need for frequent plant shut-downs by the correct selection of materials of construction and use of suitable components, e.g. combined cleaning device.

The economics involved with reheating the treated gas have led to the regenerative gas-gas heater being adopted as the state of the art by the German power industry for the reheating duty.

The same holds true in Japan, the Netherlands and Austria.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

REFERENCES

1. Frauenfeld, M. and Clay, D.E. Gewinn mit Gavo - Pilotanlage zur Wiederaufheizung des Rauchgases im Kraftwerk Wilhelmshaven. Energie. No. 12, 1980.

SOXAL^{tm} PROCESS

K.N. Mani and F.P. Chlanda

ALLIED CORPORATION
AQUATECH^{tm} SYSTEMS
7 Powder Horn Drive
Mt. Bethel, NJ 07060

The SOXAL process is a regenerative wet scrubbing system which produces a concentrated sulfur dioxide stream and is applicable for low to high sulfur coals. The sulfur dioxide can be liquified, converted to sulfuric acid or elemental sulfur. The process is based on the use of a high pH sodium solution as a scrubbing medium to remove the sulfur oxides from the flue gas. The product solution from the absorber is regenerated electrically using water splitting membranes in which th bisulfite salt is converted to the original sulfite and hydroxide that are recycled and an aqueous solution of SO_2 which is readily stripped to recover concentrated sulfur dioxide.

The SOXAL process and the associated membrane technology have been under development at Allied Corporation for over 12 years and are currently being commercialized by it's AQUATECH Systems business unit.

Process Description

A diagram of the SOXAL process is shown in Figure 1. The heart of the process is the "membrane Water Splitter" (AQUATECH cells) that electrodialytically convert salt solutions to their constituent acids and bases at low energy consumption. Key to the operation of the AQUATECH cell is the "bipolar membrane", developed by Allied Corporation.

Referring to Figure 1, SO_2-rich flue gas from the boiler is contacted with sodium sulfite solution in an absorber to substantially recover the SO_2 and convert the sulfite to bisulfite. It is assumed that almost all (99%) of the flyash particulates have been removed from the flue gas by use of electrostatic precipitators or bag houses. A pre-saturator is not necessary because any of the incoming impurities, such as chlorides, can be readily purged from the SOXAL process loop. The reactions occurring in the absorber are:

1. $2NaOH + SO_2 = Na_2SO_3 + H_2O$
2. $Na_2CO_3 + SO_2 = Na_2SO_3 + CO_2$
3. $Na_2SO_3 + H_2O + SO_2 = 2NaHSO_3$
4. $Na_2SO_3 + \frac{1}{2}O_2 = Na_2SO_4$

The reactions take place at the adiabatic saturation temperature of approximately 130°F. The first three reactions result in SO_2 absorption and the production of sodium bisulfite ($NaHSO_3$). Reaction 4 represents the oxidation of sulfite to sulfate by the oxygen present in the flue gas. A typical absorber recycle solution would have 0.5% NaOH (or soda-ash equivalent), 17% Na_2SO_3 and 7% Na_2SO_4. Typically 65-85% of the sulfite is converted to the bisulfite. Sulfate production is expected to amount to 5% of the SO_2 absorbed.

Product solution from the absorber consisting mostly of the bisulfite is filtered to remove suspended flyash impurities and fed to the AQUATECH regeneration system. A typical absorber product composition is 2.2% Na_2SO_3, 22.5% $NaHSO_3$, 7.4% Na_2SO_4.

Regeneration System

The regeneration system is composed of three parts:

(a) The two-compartment cell system (primary AQUATECH recovery unit)

(b) SO_2 stripping/recovery

(c) The three-compartment cell section (secondary AQUATECH recovery unit)

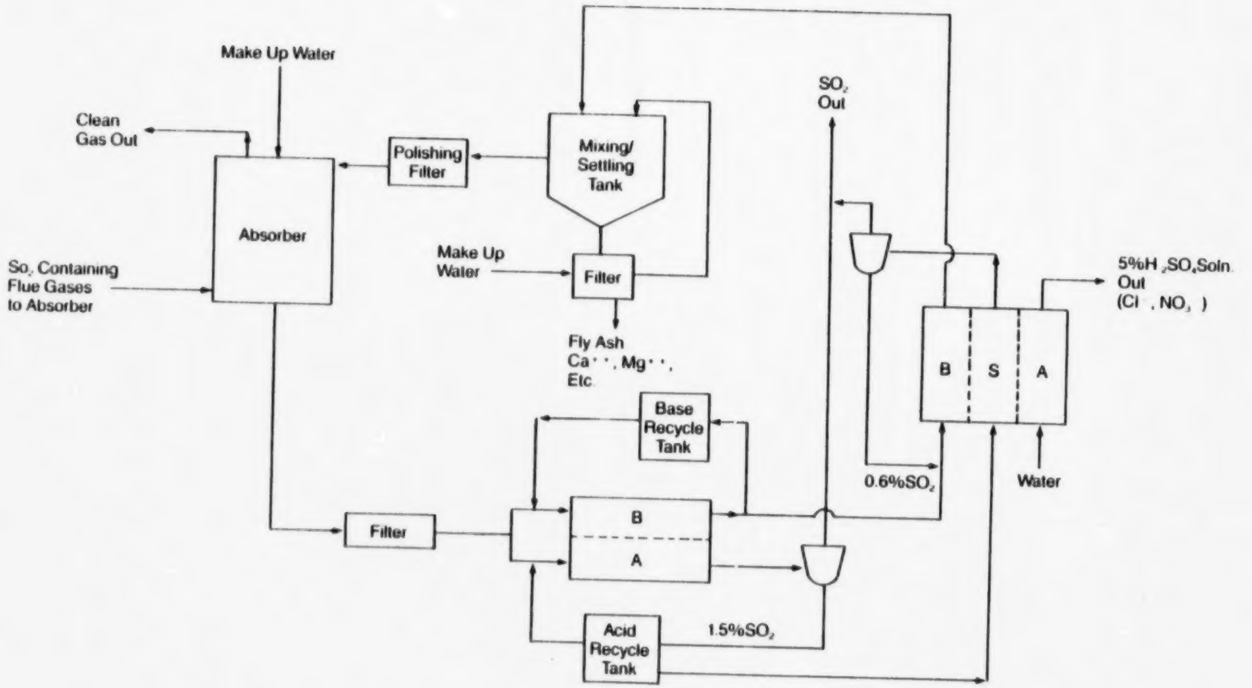

Make Up Water
Clean Gas Out
Absorber
So₂ Containing Flue Gases to Absorber
Polishing Filter
Mixing/ Settling Tank
Make Up Water
Filter
Fly Ash Ca⁺⁺, Mg⁺⁺, Etc.
Filter
Base Recycle Tank
B
A
Acid Recycle Tank
1.5%SO₂
SO₂ Out
0.6%SO₂
B
S
A
Water
5%H₂SO₄Soln. Out (Cl⁻, NO₃⁻)

FIGURE 1: SOXAL™ Process

In the regeneration section essentially all of the sodium values are regenerated as sulfite/hydroxide for recycle to the absorber. Concurrently, the SO_2 is recovered as a concentrated (95+%) stream. The small amounts of sulfate values from the oxidation step (reaction 4) along with any chlorides from the absorption step are rejected as a dilute acid stream ($HCl + H_2SO_4$). This stream is neutralized with limestone and discarded.

Bipolar Membrane

Use of the bipolar membrane is central to the regeneration process. An expanded view of its cross section is shown in Fig. 2.

The membrane consists of three regions, a cation selective region, an anion selective region and the interface between the two regions. The membrane behaves anisotropically under the influence of an electric field as is illustrated by the transport processes shown in Figure 2. Figure 2a shows that when a direct current is passed across the bipolar membrane with its cation side toward the anode and salt solution on either side, cations and anions are transported to the interface through the permselective membranes, but the passage of ions out of the interface region is limited since they would have to pass through membranes of the wrong selectivity. The salt consequently builds up at the interface and results in a low resistance in this region. When the orientation of the membrane is reversed as shown in Figure 2b and a current is passed, salt from the interface is transported to the external solutions leaving only H^+ (acid) and OH^- (base) ions from the dissociation of water to carry the current. In this mode the resistance of the interface can become high since water has a low conductivity. Our interest in bipolar membranes is in their water splitting ability, i.e. the process in Figure 2b. If the interface is made very thin, then even though it has a low conductivity the resistance of the layer can be small and the membrane may be used economically to generate acid and base.

Two Compartment Cell System

Regeneration of the spent bisulfite solution to sulfite and SO_2 is carried out in a two-compartment cell system (primary AQUATECH recovery unit). In this cell the bipolar membranes are used in conjunction with cation selective membranes. A schematic diagram of a two cell unit is shown in Figure 3. Commercially as many as 200-300 cell units are assembled to form a compact water splitting "stack". The compartments bounded between the cation membranes and the cation selective side of the bipolar are termed "acid" compartments while those between the cation membranes and the anion selective side of the bipolars are the "base" compartments.

Spent product solution from the absorber is divided into two streams and fed to the acid and base compartments of the primary AQUATECH unit. The reactions occurring in the cell are:

Acid Compartments

$$Na_2SO_3 + 2H^+ - 2Na^+ = H_2SO_3$$

$$NaHSO_3 + H^+ - Na^+ = H_2SO_3$$

$$H_2SO_3 = H_2O + SO_2$$

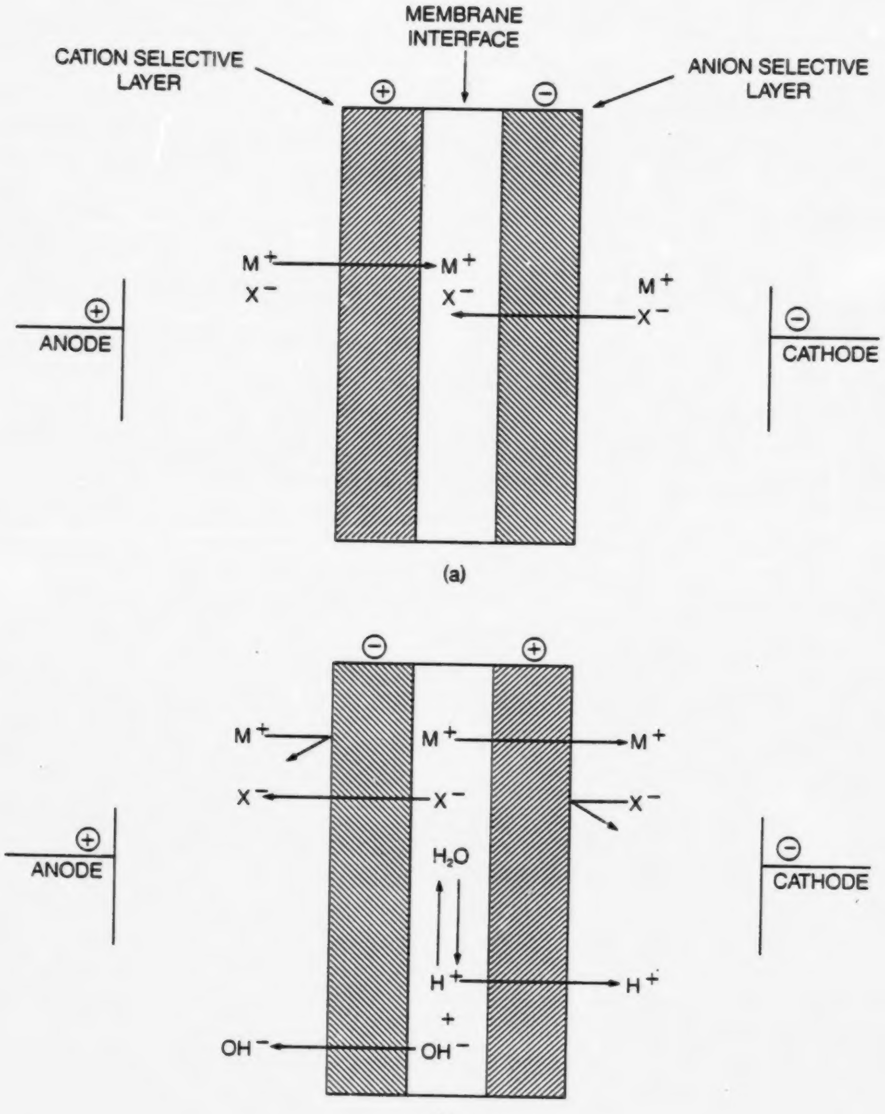
Base Compartments

$$NaHSO_3 + Na^+ + OH^- = Na_2SO_3 + H_2O$$

$$Na^+ + OH^- = NaOH$$

Thus under the influence of a direct current driving force, the sodium bisulfite is converted to SO_2 and sulfite. Any sulfite in the acid compartment feed is also converted to SO_2. In this context it is desirable to minimize the sulfite in the water splitter feed since its conversion to SO_2 requires twice as much energy as the bisulfite.

Figure 4 shows a typical plot of the overall current efficiency (defined later) as a function of bisulfite conversion. Thus the points corresponding to 0 moles/l of $NaHSO_3$ represents complete (100%) conversion of the bisulfite. When the feed solution contains no sulfate (the lower curve) the current ef-


MEMBRANE INTERFACE
CATION SELECTIVE LAYER
ANION SELECTIVE LAYER
ANODE
CATHODE
M^+
X^-
H_2O
H^+
OH^-

(a)

(b)

FIGURE 2: Bipolar Membrane Operation

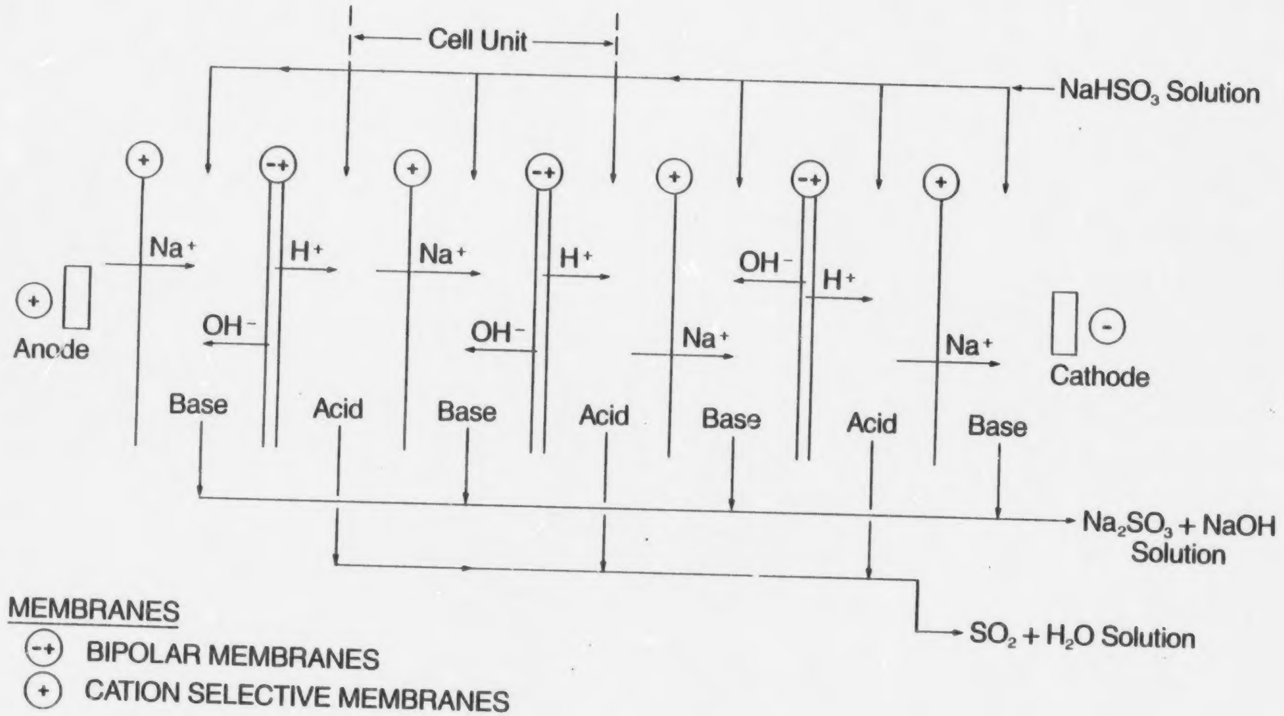

Cell Unit
NaHSO$_3$ Solution
Anode
Cathode
Na$^+$
H$^+$
OH$^-$
Base
Acid
Na$_2$SO$_3$ + NaOH Solution
SO$_2$ + H$_2$O Solution
MEMBRANES
BIPOLAR MEMBRANES
CATION SELECTIVE MEMBRANES

FIGURE 3: Two Compartment Cell System

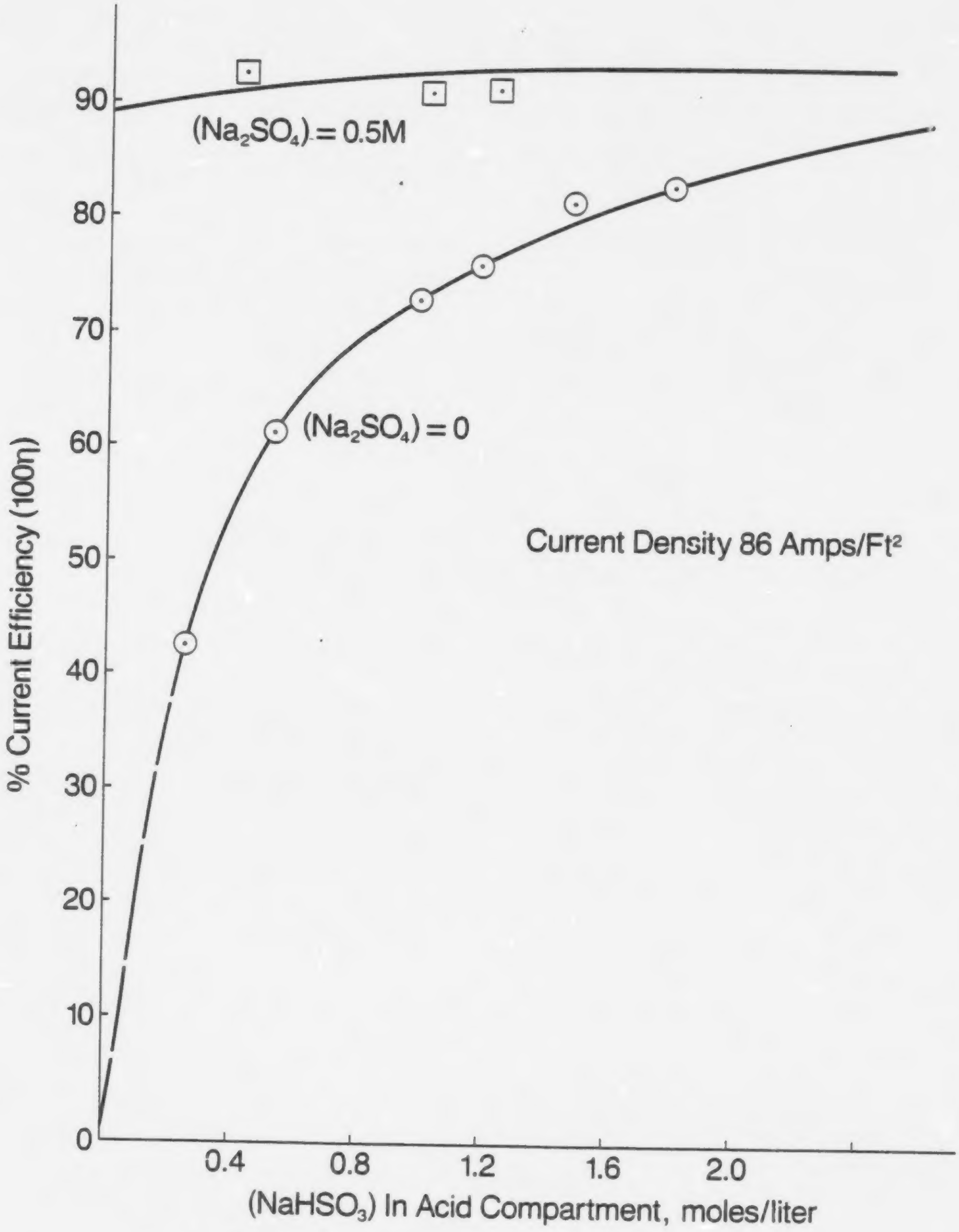


FIGURE 4: Current Efficiency As A Function Of Bisulfite Concentration

ficiency is seen to fall off rapidly as the bisulfite is converted. This is because of the competition between Na^+ and H^+ transport across the cation membranes. Only the transport of Na^+ leads to conversion of $NaHSO_3$ to H_2SO_3. As the conversion of bisulfite proceeds, the ratio Na^+/H^+ will fall to zero at complete conversion. The presence of a second source of sodium, however, solves the problem by increasing the Na^+/H^+ ratio. In the SOXAL process, since it is formed by oxidation in the absorber anyway, Na_2SO_4 is a logical choice. As can be seen from the top curve in Figure 4, the presence of 0.5M Na_2SO_4 in the bisulfite feed serves to increase the overall efficiency dramatically, over the entire conversion range.

In the two compartment cell there are two primary sources of efficiency loss, namely:

(a) H^+ transport across the cation membrane and

(b) SO_2 diffusion. The diffusion of SO_2 from the acid to base compartments results in the conversion of caustic soda/sulfite to sulfite/bisulfite.

The overall efficiency can be approximated as:

$$\eta = 1 - \eta^{c}_{H^+} - \eta^{D}_{SO_2} = \eta^{c}_{Na^+} - \eta^{D}_{SO_2}$$

Where $\eta^{c}_{Na^+}$ and $\eta^{c}_{H^+}$ are respectively the efficiencies for sodium and hydrogen ions expressed as equivalents transported across the cation membranes per faraday (96500 amp. sec./equivalent) of electricity passed. $\eta^{D}_{SO_2}$ represents the undesired inefficiency caused by the diffusion of the dissolved SO_2 from the acid compartments. η therefore gives net efficiency for generation of SO_2 in the acid compartments.

Figure 5 shows the measured efficiencies for the conversion of bisulfite to sulfite as a function of current density. As the current density is increased, the loss of efficiency caused by the diffusion of SO_2 decreases since diffusion is dependent on the concentration difference and independent of current density. At infinite current density diffusion of SO_2 is infinitesimally small and the measured efficiency is identical to the efficiency for sodium transport, i.e., $\eta = \eta_{Na^+}$. Referring to Figure 5, the top line (—·—·—) represents the sodium transport efficiency (which is not normally a function of current density.) The overall efficiency η is shown as a dashed (– – –) line. The difference between the two lines represents the efficiency loss resulting from the diffusion of SO_2.

In addition to current density two other factors determine the extent of diffusion of SO_2 namely (a) the actual concentration of SO_2 in solution, which sets the driving force for diffusion and (b) the type of membranes used.

Construction of the water splitter units is essentially modular. Each typical commercial primary recovery unit would consist of up to 300 cell units of $1 m^2$ membranes. The number of stack modules used is dependent on the amount of bisulfite to be regenerated. We estimate a 500 MW plant burning 3½%S coal (TVA basis) would require 18 operating stacks. The water splitting stacks typically operate at 110°F. All of the stacks would be operated in parallel from a common set of acid and base surge tanks. The modular nature of the process simplifies process design/scale up and affords considerable operational flexibility for the utilities.

SO_2 Stripping and Recovery

Sulfur dioxide gas is liberated from the dissociation of sulfurous acid in the acid compartment loop of the primary recovery unit. Continuous removal of the gas serves to increase overall process efficiency by decreasing its concentration in solution which in turn minimizes diffusion losses through the membranes. As much as 80-90% of the regenerated SO_2 can be recovered simply by a subatmospheric stripping operation (5-10psia). Dissolved SO_2 in the acid recirculation loop is thereby reduced to 2.5 wt.% or less which enables the AQUATECH cells to operate efficiently.

The bleed stream from the primary recovery unit is typically composed of 8-11%, Na_2SO_4, 1.5-2.5% SO_2. This stream is further stripped at a higher vacuum (∿ 2psia) to remove additional SO_2 prior to feeding (or in the salt recirculation loop of) the secondary recovery unit. Any chlorides (and nitrates) from the absorption will be present in the solution as NaCl (∿ $NaNO_3$, $NaNO_2$).

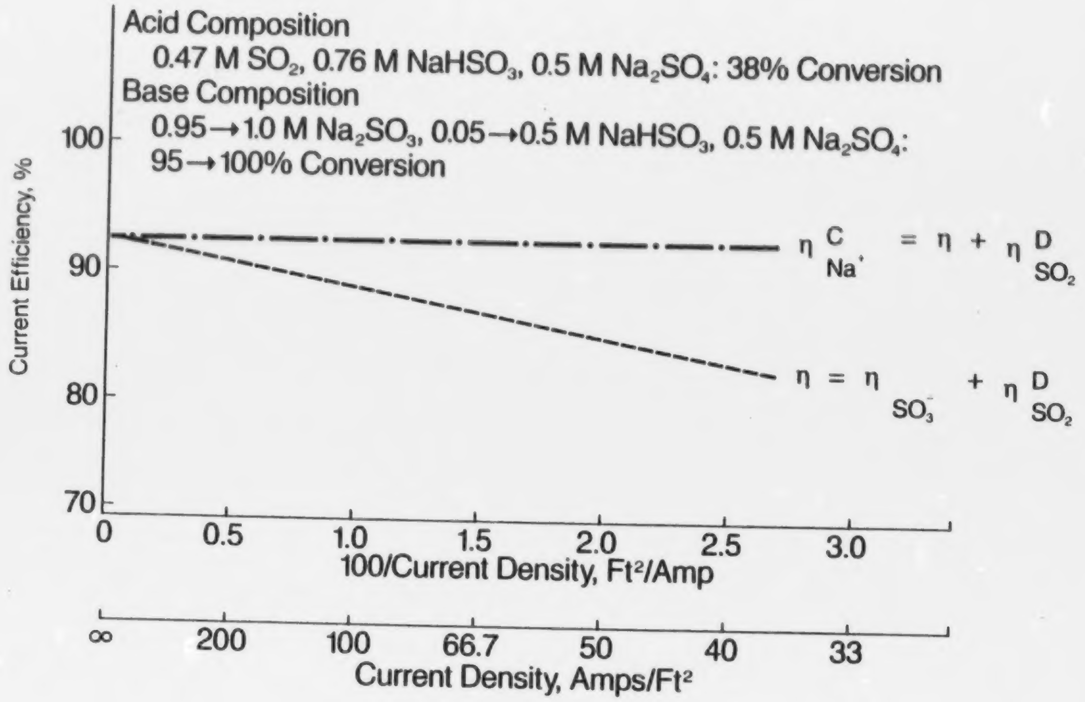


FIGURE 5: Efficiency Loss Due To Diffusion Of SO_2

Three Compartment Cell System

The three compartment cell system (Secondary AQUATECH recovery unit) is used to convert the purge sulfate to a dilute acid stream while regenerating the alkali for recycle to the absorber. Operation of the cells is shown schematically in Figure 6. The unit cell consists of a bipolar membrane, acid compartment, anion selective membrane, salt compartment, cation selective membrane and base compartment. As with the two compartment cell, a large number (200-300) of cell units would be assembled between a single set of electrodes to form an electrodialysis stack. Ion flow is as indicated in Figure 6. The reactions are:

Water Splitting $H_2O = H^+ + OH^-$

Acid Compartments $2H^+ + SO_4^= = H_2SO_4$

$H^+ + Cl^- = HCl$

Base Compartments $Na^+ + OH^- + NaHSO_3 = Na_2SO_3 + H_2O$

$Na^+ + OH^- = NaOH$

The salt compartments serve as the source of Na^+, $SO_4^=$ and Cl^- ions. Only a portion of the salt feed (equivalent to overall formation rate of sodium sulfate and chloride in the absorber) is converted in the secondary AQUATECH cells while the balance is combined with the base product stream and forwarded to the absorber.

Byproduct Recovery

Concentrated SO_2 (95wt%) is the major byproduct for the SOXAL process. It should be readily saleable after liquefaction. Alternatively, it can be converted to sulfuric acid or elemental sulfur using available technology.

A relatively small amount of acid is the second byproduct. This acid is available only as a dilute stream (5wt% is typical) and is suitable for in-house uses such as ion exchange regeneration or cleaning of process units. Alternatively, this acid stream can be neutralized with limestone and disposed of in an environmentally acceptable manner.

Process Development Status

The two compartment water splitting (Primary Recovery Unit) process has been demonstrated in a pilot plant facility at Allied's laboratories in Morristown, NJ. A flow diagram of the pilot plant is shown in Figure 7. A three stage operation was used in the test to minimize SO_2 diffusion losses and thereby increase process efficiency. (Membrane improvements since this test have eliminated the need for staging) 12" x 24" x 20 cells were used in each of the three stages. Sodium bisulfite solution was converted using the AQUATECH cells to sulfite and SO_2, recombined and reused in the test.

The pilot plant was sized to process 22 lbs/hr of SO_2 ($\sim$ 0.5 MW scale for 3.5% S coal). A summary of a 4400 hour test operation is shown in Figures 8 and 9. It is seen that the overall potential drop for the process as well as the current efficiency (i.e. product throughput) remained stable throughout the test. Bisulfite conversion was maintained at or near 100% during the test. The unit cell voltage was computed to be 1.5-1.6V, which was the design value.

It should be pointed out that even though the test was initiated with synthetic solutions, considerable amounts of impurities such as iron built up during the test. Examination of the membranes at the conclusion of the test showed them to be in excellent condition in terms of mechanical integrity and performance. There was no evidence of membrane fouling by organics or metals such as iron, calcium, etc.

The purposes of the test were the generation of performance data and establishment of membrane durability. These goals have been achieved in the pilot test. Confidence in the SOXAL process is therefore justified.

The three compartment water splitter has also been demonstrated in the laboratory as well as in field demonstrations for sulfate and fluoride salt conversions. A particularly attractive application for AQUATECH technology is the recycling of spent pickle liquors from stainless steel plants. A demonstration plant with 12" x 24" x 25 cell is presently operating successfully in this application.

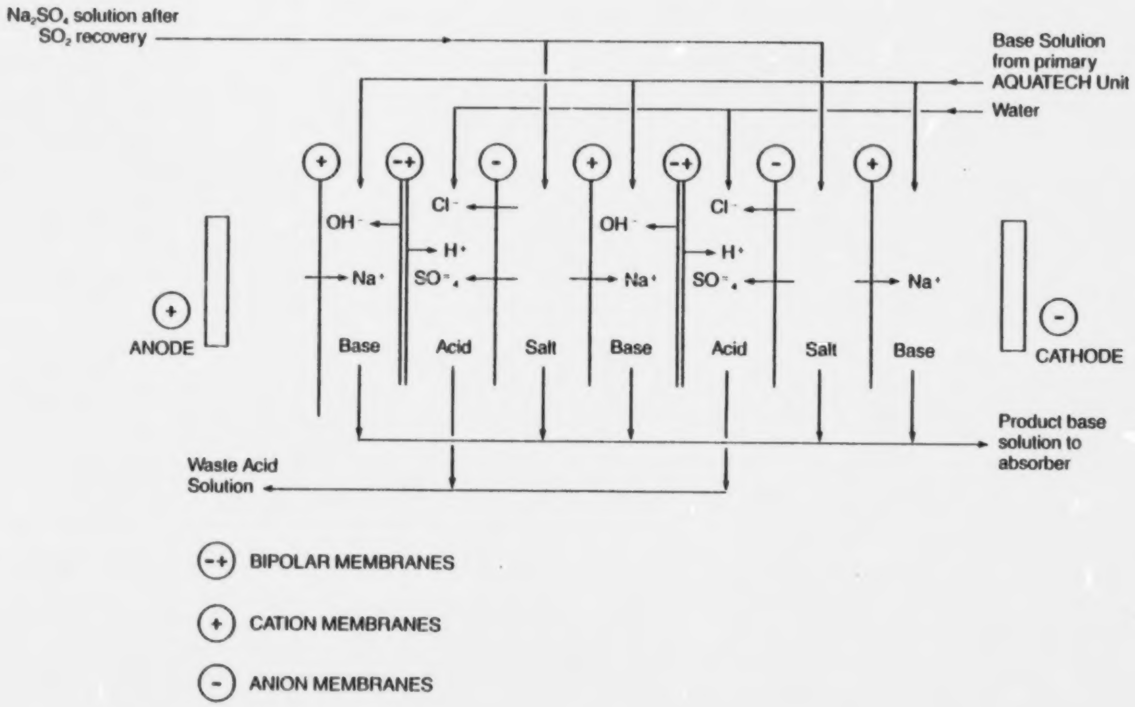
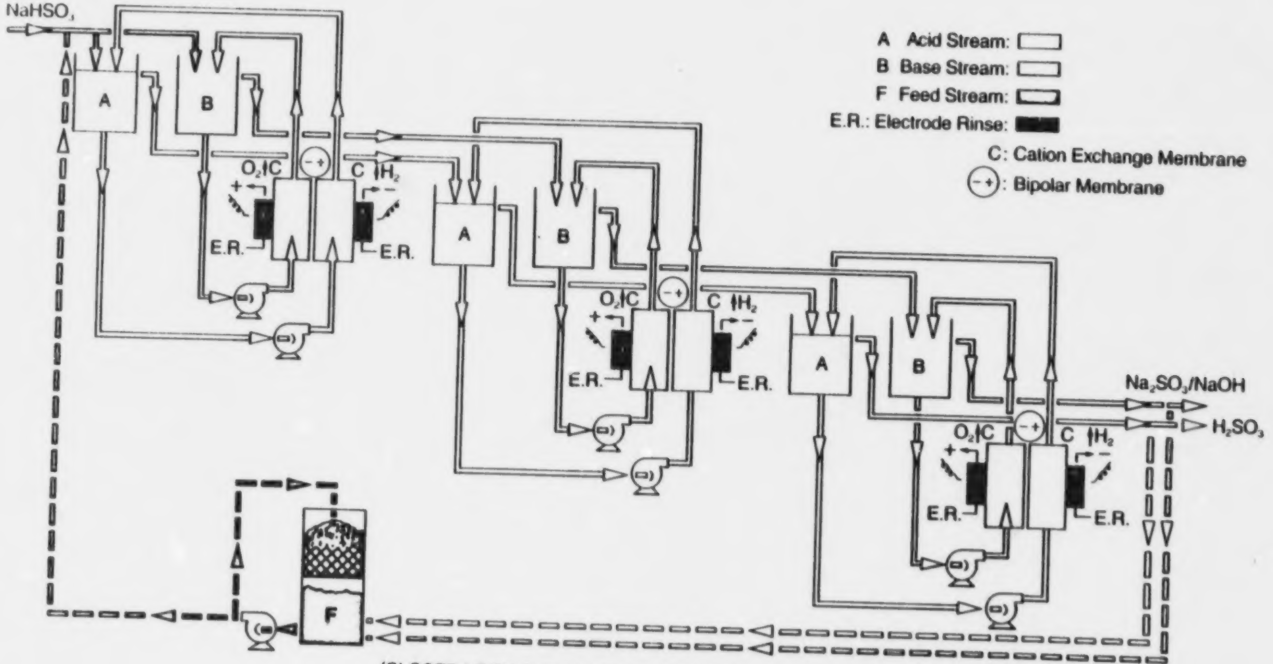

Na$_2$SO$_4$ solution after SO$_2$ recovery
Base Solution from primary AQUATECH Unit
Water
OH⁻
Cl⁻
H⁺
Na⁺
SO$_4^=$
ANODE
CATHODE
Base
Acid
Salt
Base
Acid
Salt
Base
Product base solution to absorber
Waste Acid Solution
BIPOLAR MEMBRANES
CATION MEMBRANES
ANION MEMBRANES

FIGURE 6: Three Compartment Cell System


NaHSO$_3$
A
B
O$_2$ C
C H$_2$
E.R.
E.R.
A Acid Stream:
B Base Stream:
F Feed Stream:
E.R.: Electrode Rinse:
C: Cation Exchange Membrane
(–+): Bipolar Membrane
Na$_2$SO$_3$/NaOH
H$_2$SO$_3$
F

(CLOSED-LOOP OPERATION TO REUSE ACID & BASE FOR LAB TEST)

FIGURE 7

SO_2 ED Pilot Plant Flow Diagram

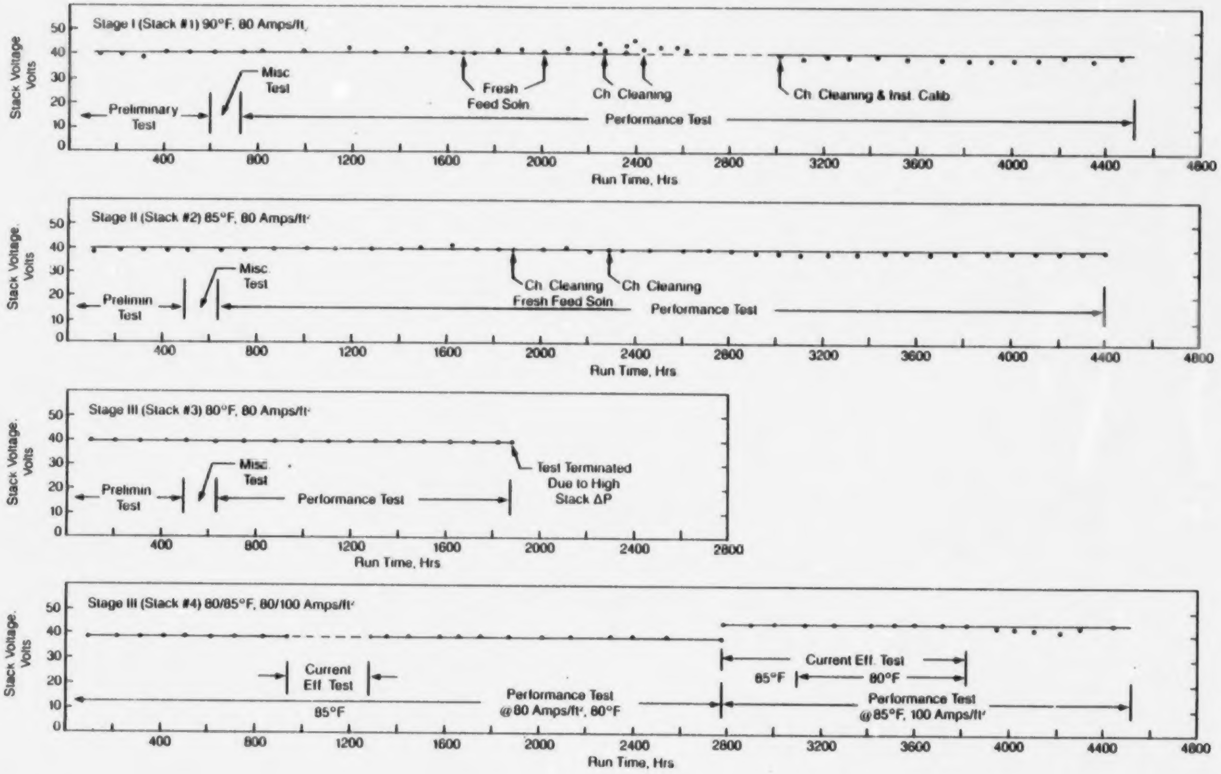


FIGURE 8: 3-Stage ED Pilot Plant Preliminary Performance Test Results

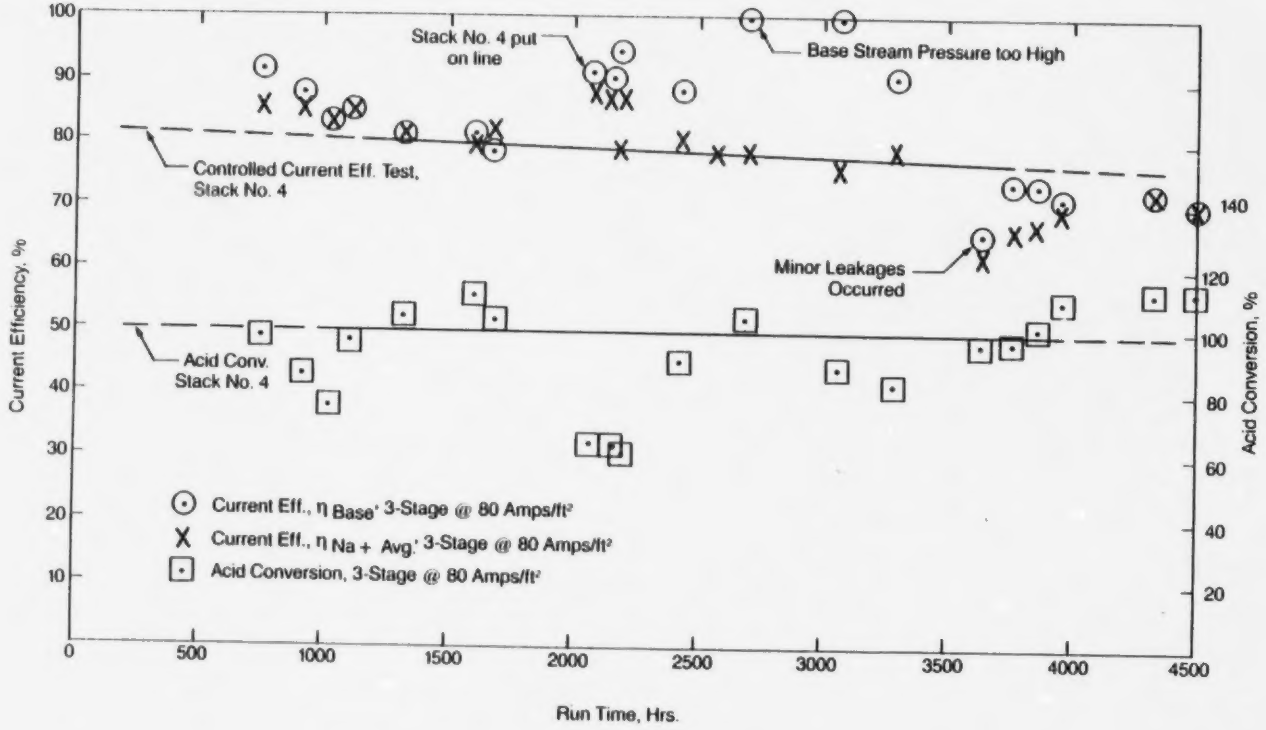


FIGURE 9: Operating Time Effect on Current Efficiency

Advantages of the SOXAL Process

Table 1 lists some of the key advantages of the process. The process is modular, making it readily amenable to scale up. The process described herein regenerates essentially all the sodium base values while recovering the bulk of the sulfur values in the form of a concentrated SO_2 stream. It is also compact which makes it attractive for use in retrofit applications and where space is at a premium.

Economic Analyses

Detailed economic studies have been carried out at Allied. The evaluation was projected for a 500MW plant burning 3½%S coal. The plant specifications (Table 2) are based on recent TVA/EPA studies. Table 3 shows projected specifications for the water splitter units. Economics of SOXAL and the other competing processes have been updated to a Dec. '84 basis.

Table 4 shows cost projections of the capital and key operating costs of regeneration for SOXAL, Wellman Lord and Mag-Ox processes. It is seen that the SOXAL process is significantly superior on both counts.

Tables 5 and 6 show our projections of the overall capital and operating costs of the SOXAL process. Two cases have been considered – one generating liquid SO_2 and the other for sulfuric acid production. Product credits ($50/ton for H_2SO_4; $72.70/ton for liquid SO_2) are based on present day acid prices. The overall operating costs estimated at 5.67 mills/kW-h. This translates to $15.1/ton of coal used by the utility.

Table 7 shows a comparison of SOXAL process with the two leading throwaway options. It is seen that while the capital requirements for the SOXAL process are somewhat higher, annual operating costs (i.e. revenue requirements) are 10-15% lower.

Conclusions

The central thrust of AQUATECH technology is in the recovery and recycle of materials used in chemical operations. The SOXAL process typifies this by recovering SO_2 which may be sold while recycling the sodium base for fluegas scrubbing. This modular membrane process has been tested extensively in Allied's laboratories and is now ready for a commercial demonstration in the 3 to 50 MW(e) scale. The process has a number of attractive features including compactness and ease of scale-up. Economic analysis shows the SOXAL process to be competitive with the throwaway processes while substantially mitigating the burden of sludge handling and disposal. Concentrated SO_2 from the SOXAL process should be readily marketable in a liquid form or could be converted to other sulfur products. Allied is eager to participate in the commercial demonstration of SOXAL technology.

AQUATECH and SOXAL are trademarks of Allied Corporation.

All statements, information and data given herein are believed to be accurate and reliable but are presented without guaranty, warranty, or responsibility of any kind, express or implied, on our part. Statements or suggestions concerning possible use of our product are made without representation or warranty that any such use is free of patent infringement and are not recommendations to infringe any patent.

TABLE 1

Features of SOXAL™ Process

- High pH Scrubbing Solution Resulting in Very High SO_2 Removal Capabilities
- Sodium-Based Clear Solution for Scrubbing
- Electrically Driven Process
- Operating Flexibility
- Built-in Load Leveling Capability
- Closed-Loop Regenerative Process
- Essentially Eliminates Hazardous Waste Streams
- Saleable Sulfur Products
- Competitive With Throwaway Processes

TABLE 2

ECONOMIC ANALYSIS BASIS

1. 500 MW Plant burning 3½% sulfur coal
2. 1.2 lb SO_2/MBTU allowable emission
3. 7200 hrs/yr
4. Dec. '84 Costs (CE. Index 330)
5. Stack Gas Reheat to 175°F via indirect steam reheat
6. Disposal pond for throwaway processes located 1 mile from power plant
7. 99+ % of flyash and particulates removed prior to FGD
8. 19750 lb/hr of SO_2 processed.
9. 988 lb/hr of SO_2 oxidized in scrubber (5% oxidation)
10. SOXAL Process projection: 67,500 tons/yr of SO_2
    5,440 tons/yr of H_2SO_4 (5 wt% soln)
11.

Process	Absorber Pressure Drop Δ P inches of H_2O	Oxidation Rate, %
Limestone Slurry	8	20
Dual Alkali	3	10
SOXAL Process	6	5

12. Cost estimates based on TVA/EPA Studies
    Reports PB80-105828 (1979)
    PB-242541 (1975)

TABLE 3

SPECIFICATIONS FOR TWO COMPARTMENT WATER SPLITTER

NaOH generated	16257.2 lb/hr = 51.255 gm equiv/sec
Current density	100A/ft^2
Cell voltage drop	1.7V
Current efficiency	92%
Effective membrane area	54,000 ft^2
Power consumption	9200 kW = 1120 kW•h/ton of NaOH

SPECIFICATIONS FOR THREE COMPARTMENT WATER SPLITTER

NaOH generated	1704 lb/hr = 5.37 gm equiv/sec
Current density	100A/ft^2
Cell voltage drop	2.0 V
Current efficiency	86%
Sulfuric acid strength	3-5 wt%
Effective membrane area	6,000 ft^2
Power consumption	1,200 kW = 1,400 kW•h/ton of NaOH

TABLE 4
REGENERATION SECTION
COMPARISON OF CAPITAL AND KEY OPERATING COSTS

Capital Investment, $M	Wellman Lord	Mag-Ox	SOXAL Process
Reagent Feed System	0.69	3.44	--
Regeneration System	23.03	14.08	12.99
Waste Handling	4.81	2.75	--
Feed Treatment	--	--	1.14
Total Direct Capital	28.53	20.27	14.13
Indirects @ 20%	5.71	4.05	2.83
Contingency @ 20%	5.71	4.05	2.83
Total Fixed Investment	39.95	28.37	19.79
Relative Capital, %	202	143	100
Key Variable Costs, $M/yr			
Steam @ $5/M BTU	5.76	--	--
Electricity @ 5¢/kW-h	0.80	0.40	3.30
Fuel Oil @ $0.90/Gallon	--	4.95	--
Soda Ash Makeup @ $125/ton	0.50	--	--
Magnesia Makeup @ $300/ton	--	2.40	--
Cell Maintenance	--	--	2.58
Total Variable Costs	7.06	7.75	5.88

TABLE 5

SOXAL PROCESS
FGD WITH BYPRODUCT RECOVERY

Products: liquid SO_2 or 93 + % H_2SO_4

	Capital, SM	
Direct Investment	Liq. SO_2	H_2SO_4
Gas Handling	5.91	5.91
SO_2 Absorption	12.71	12.71
Stack Gas Reheat	1.77	1.77
Flyash, Ca++ removal	1.00	1.00
SO_2 Regeneration	10.20	10.20
Sulfate Conversion	1.58	1.58
SO_2 Stripping	1.14	1.14
SO_2 Drying, Compression, liquefication	3.84	–
Buildings	0.70	0.70
Subtotal	38.85	35.01
Services, Utilities, Misc.	2.14	1.80
Sulfuric Acid Plant 400 T/day		6.70
Sulfuric Storage (15 days)		0.40
Total Direct Investment (TDI)	40.99	43.91
Indirect Investment		
Indirects @ 20%	8.20	8.78
Contingency @ 20%	8.20	8.78
Total Fixed Investment	57.39	61.47
Other Capital Charges		
Allowance for Startup & Modifications	4.92	5.31
Interest During Construction	7.38	7.90
Total Depreciable Investment	69.69	74.68
Land @ S4800/acre	0.01	0.01
Working Capital	1.64	1.78
Total Capital Investment, SM (TCI)	71.34	76.47
S/kW	S143	S153

TABLE 6

SOXAL PROCESS
FGD (WITH BYPRODUCT RECOVERY)

Operating Costs
Regulated Utility Economics
7200 hrs/yr

Direct Costs	Annual Quantity	Unit Cost	Total Cost, $M Liq. SO_2 Product	Total Cost, $M H_2SO_4 Product
Raw Materials:				
NaOH Makeup, tons	80	$ 250/ton	0.02	0.02
Cell Maintenance	---	---	2.58	2.58
Conversion Costs:				
Operating Labor & Supervision, man hrs	35,480	$ 18/hr	0.64	0.64
Utilities:				
Steam, MBTU	503,800 (for SO_2) 345,400 (for H_2SO_4)	$ 5/M BTU	2.52	1.73
Process Water	125,000 kgal	$0.25/kgal	0.03	0.03
Electricity				
ED cell section	66,000,000 kW•h	$0.05/kW•h	3.30	3.30
Others	23,100,000 kW•h	$0.05/kW•h	1.16	1.16
Maintenance:				
4% of TDI (for non ED cell areas)			2.32	2.52
Total Direct Costs			12.57	11.98
Indirect Costs				
Capital Charges:				
Depreciation	6% TDI		4.18	4.48
Cost of Capital	10% TCI		7.14	7.65
Overheads:	2% of Capital		1.43	1.53
Total Operating Costs, $M			25.32	25.64
Product Credits				
Liquid SO_2	67,500	$72.3/ton	(4.88)	0
Sulfuric Acid	104,000	$ 50/ton	0	(5.20)
5 wt% H_2SO_4 solution, tons	5440	0	0	0
Net Operating Cost, $M			20.44	20.44
Unit Operating Cost, mills/kW•h			5.67	5.67

TABLE 7

SUMMARY OF CAPITAL AND OPERATING COSTS

Capital Costs, $M	Limestone Slurry Process (On-Site Waste Disposal)	Dual Alkali (On-Site Waste Disposal)	SOXAL Process LIQ SO_2	SOXAL Process H_2SO_4
Total Direct Investment	36.11	37.67	40.99	43.91
Indirect Investment	14.44	15.06	16.40	17.56
Total Fixed Investment	50.55	52.73	57.39	61.47
Other Capital Charges	10.83	11.30	12.30	13.21
Total Depreciable Investment	61.38	64.03	71.34	76.47
Land Costs	1.46	1.21	0.01	0.01
Working Capital	1.44	1.51	1.64	1.78
Total Capital Investment	64.28	66.75	71.34	76.47
Total Capital, $kW	130	134	143	153
Operating Costs, $M				
Direct Operating Costs	10.87	12.31	12.57	11.98
Indirect Operating Costs	11.34	11.80	12.75	13.66
Total Operating Costs	22.21	24.11	25.32	25.64
Byproduct Credits ($72.70/ton of SO_2, $50/ton of H_2SO_4 $0/ton for dilute H_2SO_4 soln.)	(0.0)	(0.0)	(4.88)	(5.20)
Net Operating Cost, $M	22.21	24.11	20.44	20.44
Unit Operating Cost, mills/kW•h	6.20	6.70	5.67	5.67

Session 9: SPRAY DRYER FGD

Theodore G. Brna, Chairman
U.S. Environmental Protection Agency
Research Triangle Park, NC

SPRAY DRYER FGD EXPERIENCE

JOY-NIRO INSTALLATIONS

J.R. Donnelly - A/S Niro Atomizer
S. Wilson - Joy Manufacturing Company
L. P. Matis - Northern States Power Company
R. Eriksen - Basin Electric Power Cooperative
R.D. Emerson - Sunflower Electric Cooperative
J.C. Fooks - Platte River Power Authority

ABSTRACT

Joy-Niro lime based Spray Dryer Absorption (SDA) Flue Gas Desulfurization systems are in commercial operation on the following coal-fired utility boilers:

Riverside Generating Station - Units 6 & 7
Antelope Valley Station - Unit 1
Holcomb Generating Station - Unit 1
Rawhide Generating Station - Unit 1

More than nine years of operating experience have been accumulated for these stations. These systems employ spray dryer absorbers, equipped with single rotary atomizers, and reverse air bagfilters for SO_2 and particulate removal. The systems range in size from 600,000 ACFM (110 MW) to 2,100,000 ACFM (450 MW) and treat flue gases from the firing of North Dakota lignite, Powder River Basin subbituminous coal, and a blend of subbituminous coal and petroleum coke. Each system incorporates unique site specific design features for the flue gas cleaning systems and for waste conditioning/disposal.

This paper presents a comparison of the plant designs and performance. Compliance/Performance test results are compared to design values. Start-up problems are reviewed and current plant reliability is discussed. Waste conditioning and disposal practices at each site are described and results of physical/chemical tests of the endproduct are presented.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

INTRODUCTION

Spray Dryer Absorption (SDA) is a relatively new technology for Flue Gas Desulfurization (FGD). Development and application of this technology have occured quite rapidly since the award of the first full-scale utility SDA system in 1978. Today more than 9000 MW worldwide have been committed to SDA-FGD. The rapid acceptance of this technology is in part due to extensive pilot plant programs conducted at several generating stations between 1977 and 1980. These programs have shown that SDA offers a simpler and less costly alternative to lime/limestone wet scrubbing.

Joy Manufacturing Company and Niro Atomizer Inc. (Joy/Niro) are a market leader in the SDA field. Since their first award in 1978 (Basin Electric Power Cooperative's Antelope Valley Station No. 1) Joy/Niro have received 13 utility orders representing more than 5,000 MW of generating capacity. A/S Niro Atomizer has received orders for additional 1700 MW of generating capacity outside the U.S. Four Joy/Niro systems are in commercial operation and have accumulated more than 9 years of operational experience.

The first full scale utility SDA system was started up in December 1980 treating flue gas from units 6 and 7 of Northern States Power Company's (NSP) Riverside Generating Station. This system was supplied as a demonstration plant and more than two years of extensive testwork has been conducted by Joy/Niro and NSP to characterize the SDA process. NSP took full responsibility for the operation of the system in January 1983. They have continued to operate the plant commercially and have participated in several test series to further characterize the process.

In May, 1983 the first lime based multichamber SDA system started up on unit 1 of Basin Electric Power Cooperative's Antelope Valley Station. This system represents one of the largest operating SDA systems treating more than 2.1 million acfm of flue gas. The system successfully completed compliance testing in August, 1983, and performance testing in October, 1983.

The 319 MW Holcomb no. 1 unit and the 275 MW Rawhide no. 1 unit both fire a Powder River Basin subbituninous coal and are designed for 80 percent SO_2 removal. These units are very similar in design and started up in May 1983 and January 1984 respectively. Compliance tests have been completed, and the systems are in commercial operation.

The four systems operate with high availabilities (90-99 percent), however, several problem areas have required resolution by equipment, controls, or operational modifications. Some problems areas are still being resolved. No generic problems have been found with the SDA process or the large spray dryers.

This paper presents a description of these four SDA systems and results of compliance/performance testing. Start-up problems are discussed and current plant availability is reviewed. Waste conditioning and disposal practice at each site are described. Results of physical-chemical tests of laboratory and on-site conditioned endproduct are presented. Comparisons are made of physical-chemical properties for core samples from the Holcomb landfill site with laboratory prepared Proctor cylinders.

PLANT DESCRIPTIONS

These plants which represent the first generation of a new technology have been described in detail in the literature (1-5, 12, 13, 14) and will be briefly discussed here. The systems incorporate common design features based on Joy/Niro's extensive pilot plant research programs and Niro's experience in spray drying. The features include:

- Large spray dryers equipped with a single rotary atomizer
- Compound gas disperser design
- Two point discharge
- Patented recycle system (6)
- High feed slurry solids content

General plant design data are presented in Table 1 and flue gas cleaning system parameters in Table 2.

TABLE 1. PLANT DESIGN DATA

	RIVERSIDE No. 6 & 7	ANTELOPE VALLEY No. 1	HOLCOMB No. 1	RAWHIDE No. 1
Rating MW	110 (Gross)	440 (Gross)	319 (Gross)	250 (Net)
Owner	Northern States Power Company (NSP)	Basin Electric Power Cooperative (BEC)	Sunflower Electric Cooperative, Inc. (SEC)	Platte River Power Authority (PRPA)
Location	Minneapolis, Minnesota	Beulah, North Dakota	Holcomb, Kansas	Wellington, Colorado
Architect/Engineer	Black & Veatch	Stearns-Roger	United Eng. & Constructors	Black & Veatch
Boiler type	B&W - Wall fired	C.E. - Tangential	B&W - Opposed wall	C.E. - Tangential
Fuel				
Type	Coal-coke blend Subbit - Sarpy Creek	Lignite	Subbit, Powder River Basin	Subbit, Powder River Basin
HHV Btu/lb	9300	6600	8513	8800
S %	1.15 to 3.5	0.68	0.34	0.34
Ash %	7.5	7.8	5.6	4.7
Moisture %	23.9	37.1	29.7	28.0
Gas Flow acfm	645,000 at 360°F	2,055,000 at 310°F	1,340,000 at 253°F	1,356,000 at 276°F
Start-up	SDA - Dec. 80 Boilers - 1949	May 83	May 83	Dec. 83
Compliance testing	N.A.	Aug. 83	March 84	Sept. 84
Performance testing	N.A.	Sept. 83	April 84	-

The capacities of the units range from 110 MW to 450 MW and flue gas flowrates at the inlet of the SDA system range from 645,000 to 2,055,000 acfm.

Fuels fired include; a blend of Sarpy Creek subbituminous coal (85 %) and petroleum coke (15 %), North Dakota Lignite, and Power River Basin subbituminous coal. Normal sulfur contents range from 0.34 to 1.15 percent, however, 3.5 percent sulfur Illinois coal has been fired at Riverside several times for extended periods for test purposes. Antelope Valley Unit 1 and Holcomb Unit No. 1 have completed compliance and performance testing. Rawhide Unit 1 has completed compliance testing.

TABLE 2. FGC SYSTEM DESIGN PARAMETERS

		RIVERSIDE No. 6 & 7		ANTELOPE VALLEY No. 1	HOLCOMB No. 1	RAWHIDE No. 1
SPRAY DRYERS						
Number	operating + spare	1		4 + 1	2 + 1	2 + 1
Dia. x Cylindrical height	ftxft	46x36		46 x 36	50.9 x 39.4	46 x 36
Gas residence time	sec	10		10	10.6	10
Atomizer	HP	700		700	700	800
Exit temp.	oF	153		168	179	162
Approach to adsat.	oF	18		35	50	35
Feed solids	%	30 - 50		35	40-45	45
BAGHOUSE						
Number		1		2	2	2
Compartments, total		12		28	28	24
Bags/Compartment		250		288	286	274
Air/Cloth ratio, gross		1.07*		2.19	1.55	1.77
net		2.00		2.47	1.93	2.30
Inlet temp.	oF	153		185 (ID Fan)	179	170
Bagmaterial		Fiberglass & others		Fiberglass Teflon coated	Fiberglass Teflon coated	Fiberglass Teflon coated
Dia. x Length	inxft	12 x 35		12 x 35	12 x 32.6	12 x 34
Pressure drop	in, WG	4.0		4.0	3.5	3.5
SLAKERS						
Type		Ballmill	Detention	Ballmill	Ballmill	Ballmill
Capacity	TPH	1	5	2 x 6	2 x 2.6	2 x 3
Reheat		NO		Hot gas bypass	No	Warm gas bypass
PERFORMANCE - GUARANTEES		Low S	High S			
SO_2 removal	%	70**	90**	62 ave, 78 max	80	80
SO_2 inlet - design	ppm			800 ave, 1380 max	1000	500
SO_2 emissions	lb/10^6 Btu			0.78	0.48	0.13
Particulate emissions	lb/10^6 Btu			0.04	0.03	0.03
Stoich. ratio	Ca/SO_2 - in			0.54 - 1.45	1.10	1.15
Power consumption	kW			2257	2271	2349
Pressure drop	in, WG			13.4	14.6	16.9

* Sized for non SDA operation - A/C reflects lower acfm due to SDA

** Design values

The SDA system design operating conditions were selected based on economic evaluations. The final design conditions resulted in the lowest evaluated costs considering site specific costs for lime, power consumption, pressure drop, and reheat. These are reflected in the differences in feed solids, spray dryer outlet temperatures, and the use of by-pass reheat.

The SDA systems all include large spray dryers (46-50 ft. dia.) equipped with compound gas dispersers and a single 700 or 800 horsepower rotary atomizer. The spray dryers incorporate a two point discharge and are sized for approximately 10 seconds (gross) gas residence time. Design spray dryer exit temperatures range from 18 to 50°F above the adiabatic saturation temperature. The baghouses are all reverse air cleaning type with net air to cloth ratios between 1.93:1 and 2.47:1. The selection of fiberglass bags (10 oz. fabric) with a 10% Teflon finish was based on an evaluation of a number of fabrics and finishes tested during the initial operation of the Riverside demonstration plant.

Each system employs a ball mill slaker to hydrate the raw lime and grind up any grits. All systems include a spare spray dryer (except for Riverside), spare ball mill slaker, and spare pumps for all slurry service.

Required SO_2 removal efficiencies range from 70 to 89 percent with testing at Riverside going upto 95 percent removal for both low and high sulfur fuels. Overall system pressure drops range from about 10 to 16.8 inches water gauge. All systems include a Honeywell TDC 2000 control system.

Figure 1 shows the gas flow path for the Holcomb Station and is used to illustrate the gas flow path in general. The flue gas after leaving the economizer section of the boiler passes through the air preheater and is ducted to the inlet of the flue gas cleaning system.

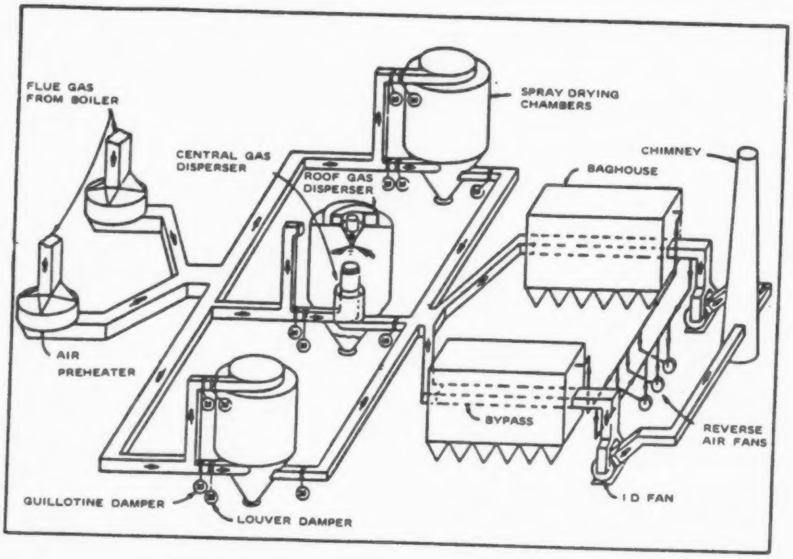


Figure 1. Holcomb Station - Gas flow sheet

Here the flue gas flow is split between the operating chambers. The flue gas stream is again split prior to entering the spray dryer such that approximately 60 percent of the gas enters the chamber through the roof gas disperser and 40 percent through the central gas disperser. Figure 2 shows in detail the spray dryer compound gas disperser. Adjustment of the two flows and positioning of the gas disperser vanes control the shape of the cloud of atomized droplet and provides efficient mixing of the flue gas and absorbent in a narrow zone around the atomizer.

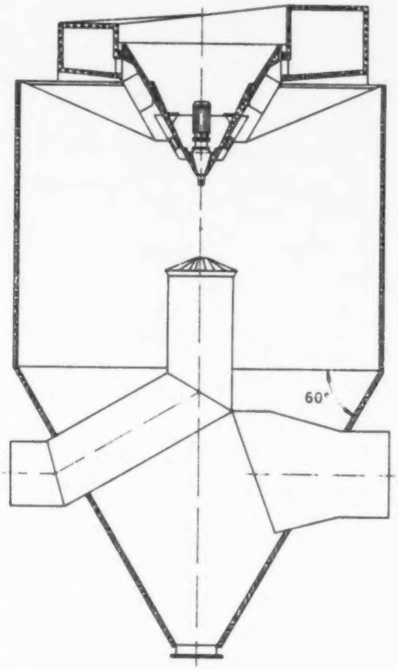


Figure 2. Spray dryer compound gas disperser

In this zone the reagent feed slurry simultaneously dries and reacts with acid gases (SO_2, SO_3, HCl) forming dried particles of spent reagent. The flue gas is at the same time cooled to between 18 and 50°F above the adiabatic saturation temperature. Typically 60 to 70 percent of the total SO_2 absorption takes place in the spray dryer.

The design of the gas dispersers and outlet duct imparts a cyclonic action to the flue gas and results in the drop out of 5 to 10 percent of the dry reagent. The flue gas containing approximately 90 percent of the dried solids after leaving the spray dryer is ducted to the reverse air type baghouse for particulate collection and additional SO_2 removal. A portion of hot flue gas (from before or after the air preheater) is sometimes bypassed around the spray dryer to reheat the flue gas before the baghouse. This is done for corrosion protection and to raise the stack exit temperatures.

Figure 3 is a simplified flow diagram showing the feed preparation system for the Antelope Valley Unit No. 1 SDA system. The feed preparation system designs for Antelope and Riverside are very similar. Holcomb and Rawhide are slightly different.

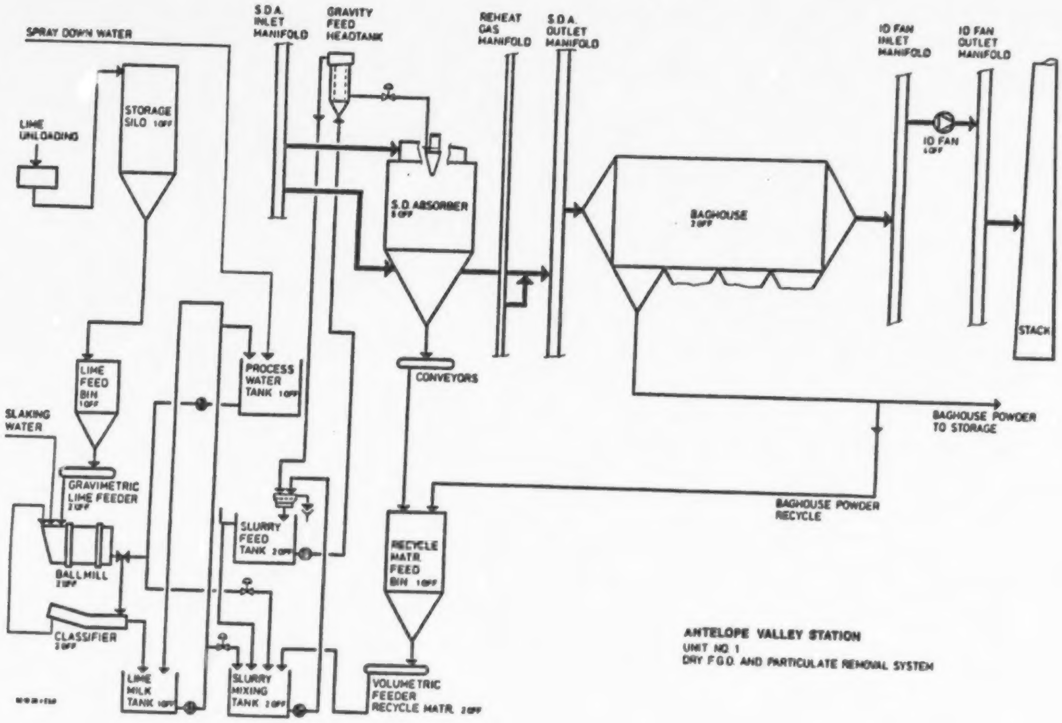

SPRAY DOWN WATER
S.D.A. INLET MANIFOLD
GRAVITY FEED HEADTANK
REHEAT GAS MANIFOLD
S.D.A. OUTLET MANIFOLD
ID FAN INLET MANIFOLD
ID FAN OUTLET MANIFOLD
LIME UNLOADING
STORAGE SILO 1OFF
S.D. ABSORBER 3OFF
BAGHOUSE 2OFF
ID FAN 4OFF
STACK
CONVEYORS
LIME FEED BIN 1OFF
PROCESS WATER TANK 1OFF
SLAKING WATER
GRAVIMETRIC LIME FEEDER 2OFF
BALLMILL 2OFF
CLASSIFIER 2OFF
SLURRY FEED TANK 2OFF
RECYCLE MATR. FEED BIN 1OFF
BAGHOUSE POWDER TO STORAGE
BAGHOUSE POWDER RECYCLE
LIME MILK TANK 1OFF
SLURRY MIXING TANK 2OFF
VOLUMETRIC FEEDER RECYCLE MATR. 2OFF
ANTELOPE VALLEY STATION
UNIT NO. 1
DRY F.G.D. AND PARTICULATE REMOVAL SYSTEM

Figure 3. Antelope Valley Station - Flow diagram

Pebble lime is fed from the lime feed bin to the ball mill slaker via a weigh belt. In the slaker the quicklime hydrates and any grit present is ground up. The lime flows to the lime milk tank from which it is pumped to the slurry mixing tank. A small portion of the lime can also be added to the process water tank for softening.

In the slurry mixing tank, water and recycled product are mixed together to form an alkaline feed slurry containing from 35 to 50 percent solids. The lime content of the slurry is controlled by either the outlet SO_2 concentration or the SO_2 removal. The feed solids content is selected based on spray dryer outlet temperature, flyash and product alkalinity, and slurry viscosity constraints.

From the mixing tank the slurry is pumped to the feed tank via a screen to remove any oversized particles. The feed slurry is then pumped to the head tanks where approximately half of the flow overflows back to the feed tank via a screen. Another screen is provided in the head tank to ensure that no oversized particles reach the atomizer. From the head tank the feed slurry flows by gravity to the atomizer. Control of feed flow is based on the spray dryer outlet temperature.

Product is collected from both the bottom of the spray dryer and from the baghouse hoppers. Mechanical conveyors are used under the spray dryer and mechanical or pneumatic under the baghouse. The collected product is transported either to the recycle material feed bin or to the waste disposal silo.

Holcomb and Rawhide employ slightly different feed preparation systems. In these systems the recycle material is mixed with water in a recycle mix tank to form a slurry of approximately 45-50 percent solids. The recycle slurry is then pumped via a screen to the feed tank. Here lime milk, head tank return and water are added, also through the screen, to form a feed slurry of the needed alkalinity and solids content. The feed slurry is then pumped to the head tank or to an intermediate feed tank. This system offers a faster response to changes in boiler flue gas conditions.

The Riverside SDA system includes a single 46 foot diameter spray dryer, F600 atomizer, and baghouse. The system was designed as a demonstration plant and incorporates several features for testing such as a reagent preparation system including both ball mill and detention slakers. A Honeywell TDC 2000 computer control system was included because of its capabilities in system control, data logging and correlation, parameter trending and report preparation to support the demonstration test program.

Antelope Valley Unit No. 1 SDA system includes five 46 foot dia spray dryers (4+1 spare) and two 14 compartment baghouses. The spray dryers operate at a 35°F approach to adiabatic saturation temperature. Reheat is provided between the spray dryers and the baghouses by mixing in hot flue gas taken from before the air preheater, however, reheat is not normally used.

Holcomb Unit No. 1 includes three 50.9 foot diameter spray dryers (2+1 spare) and two 14 compartment baghouses. These spray dryers are the largest operating spray dryers in the world. The spray dryers operate at a 50°F approach temperature and reheat is not included.

Rawhide Unit No. 1 also includes three 46 foot diameter spray dryers and two 12 compartment bagbouses. The system is designed to operate at a spray dryer outlet temperature of 35°F above the adiabatic saturation temperature with reheat by bypassing flue gas from after the air preheater to raise the baghouse inlet temperature to 170°F.

OPERATION & TEST RESULTS

All of the plants have been in operation for more than one year. The Antelope Valley, Holcomb, and Rawhide stations have completed compliance testing. Riverside has been extensively tested by Joy/Niro, NSP and independent contractors (3, 5, 13).

Operating data for each plant are presented in Table 3. Results of independent tests of Riverside and guarantee tests for Antelope and Holcomb are given in Table 4.

TABLE 3 OPERATING DATA

		RIVERSIDE No. 6 & 7	ANTELOPE VALLEY No. 1	HOLCOMB No. 1	RAWHIDE No. 1
Type data		Operating	Compliance	Compliance	Compliance
Unit load, gross	MW	40 - 100	468	337	267
Stack gas flow	acfm	250 - 600 x 10^3	1,880,000	1,227,667	1,115,000
Stack gas temp.	°F	148 - 158	222	192	189
Approach to adsat temp.	°F	25	35	50	50
Feed solids	%	35	34	45	45
SO_2					
Inlet	ppm	900	668	285 (est.)	252
Outlet	ppm	125	171	65 (est.)	68
SO_2 removal	%	80 - 95	74	76.3*	72
SO_2 emissions	lb/10^6 Btu	0.9	0.46	0.18	0.186
Particulate emissions	lbs/10^6 Btu	0.01	0.015	0.019	0.0076
Opacity		< 5	N.A.	< 5	3

* 30 day rolling average

The Riverside units operate as peaking units and as such go through many start-ups and load changes each month. The units are typically banked over the week-end and restarted each Monday. The SDA is typically operated at a 25°F approach temperature with 35 percent feed solids. Lime addition is controlled to achieve about 125 ppm SO_2 in the stack. This corresponds to 80-85 percent removal and maintains SO_2 emissions below 0.9 lbs/10^6 BTu.

Antelope Valley Unit No. 1 is base loaded and operates at or above its rated capacity. Holcomb Unit No. 1 is a load following unit with a capacity factor of between 60 and 70 percent depending on the time of the year. Compliance tests results for SO_2 removal were 74 and 76.3 percent for Antelope and Holcomb respectively. This corresponds to SO_2 emission levels of 0.46 and 0.18 lbs/10^6 BTu. Particulate emissions were between one half and two thirds the permitted value and opacity reading were five percent or less. For Antelope Valley No. 1 the required SO_2 removal efficiency will increase to 81 percent for average sulfur coal and 89 percent for maximum sulfur coal when Antelope Valley Station Unit No. 2 begins commercial operation in July 1986.

Rawhide Unit No. 1 operates at an approach temperature of approximately 35°F with 45 percent feed solids. Inlet SO_2 concentrations are typically 260 ppm and the SO_2 removal efficiency is maintained at about 82 %. Compliance tests were conducted in September 1984.

TABLE 4 TEST RESULTS

		RIVERSIDE No. 6 & 7*		Antelope Valley No. 1		Holcomb No. 1	
		Low sulfur	High sulfur	Tested	Guarantee	Tested	Guarantee
Unit load, gross	MW	65	65	463	435 **	327	319 **
Coal sulfur	%	1.60	3.15	0.68 **		0.38	1.0 **
Flue gas flowrate, SDA inlet	acfm	429,000	412,000	2,040,000	2,055,000 **	1,498,333	1,313,000 **
Flue gas temp., SDA inlet	°F	329	322	314	310	260	253
Chambers in service		1	1	4	4	2	2
SDA outlet temperature	°F	145	144	166		177	
Approach to adsat. temp.	°F	20	18	33	35 **	51.6	50 **
Baghouse outlet temp.	°F	145	144	189	185	184	
SO_2							
Inlet to SDA	ppm	810	1972	655	800 **	320	1000 **
Stack	ppm			220	304 **	61	200**
Removal	%	74.3	88.1	63.9	62	81	80
Particulate							
Emissions	gr/dscf	0.002	0.004	0.0025	0.012	0.0051	0.009
	lbs/10^6 Btu	0.004	0.009	0.005	0.04	0.01	0.03
Opacity				5	20	4 - 5	20
Stoich.ratio		0.71	1.33 (0.98)***	0.47	0.54	0.89	1.10
System pressure drop	in, WG	7.2	6.9	14.96	13.4	10.7 ****	14.6
System power consumption	kW	-	-	2238	2757	3160	2271
Baghouse air inleakage	%					2.0	0.7
Feed solids		29	30	41	35**	45	50**

* Data taken from EPRI field evaluation tests 1983
** Design value
*** S.R. for chloride enhanced tests
**** With 3 chambers in service

Results of Joy/Niro and NSP testing at Riverside have been previously reported (3, 5). The test results presented are from an EPRI-EPA sponsored field evaluation of a utility dry scrubbing system conducted by the Radian Corporation from July to October, 1983 (13). The test program included testing of both low and high sulfur coal at nominal 75 and 90 percent removal respectively. The low sulfur coal contained 1.60 percent sulfur and the high sulfur coal 3.15 percent. The stoichiometric ratios were 0.71 and 1.33 for 74.3 percent removal for the low sulfur coal and 88.1 percent removal for the high sulfur coal. For high sulfur coal the positive effects of chloride addition were also evaluated. At 89 percent removal the stoich. ratio was lowered from 1.33 to 0.98. These results are in good agreement with earlier testing performed by Joy/Niro (3, 5).

Both the Antelope Valley and Holcomb stations have successfully completed performance guarantee testing. Performance testing at the Rawhide station has not been conducted.

Antelope Valley station guarantee performance testing took place in September and October, 1983. Guarantee parameters measured were SO_2 removal, stoichiometric ratio, particulate emissions, power consumption, pressure drop, and boiler heat loss due to the use of the hot gas bypass. Four SDA's were placed in service at a boiler load of 463 MW corresponding to a design flue gas flow of 2,055,000 acfm at the SDA inlet. Two baghouse compartments were isolated and two compartments were in cleaning during the testing. Normal soot blowing and ash handling took place during the tests.

Results of the tests show the system meeting guarantees for SO_2 removal, stoichiometric ratio, particulate emissions, power consumption, and boiler heat loss. The system pressure drop exceeded the guarantee value by about 1.5 inches WG. This is attributed to higher than expected pressure drop for the SDA inlet manifold and the baghouses. The SDA outlet ducts have since been modified to reduce the pressure drop.

The performance guarantees included a debit/credit formula based on present worth costs for each parameter guaranteed. The sum of these debits and credits serve as the basis for determining overall compliance with performance guarantees. For Antelope the penalty for the high system pressure drop was more than offset by lower than guarantee values for stoichiometric ratio and power consumption. Based on these results Basin Electric and Joy/Niro proceeded with contract close out.

Holcomb Unit No. 1 performance tests were conducted in April, 1984. The unit operated at full load with two SDA's in service, two baghouse compartments isolated and two compartments in cleaning . The gasflow rate measured at the system inlet was approximately 14 percent greater than the design. Guarantees included SO_2 removal, stoichiometry, particulate emissions, system pressure drop, power consumption, baghouse inleakage, compressed air consumption, and waste product properties.

Results of the test showed that the SDA system met performance guarantees for SO_2 removal, stoichiometric ratio, particulate emissions, pressure drop, air inleakage, and compressed air consumption. The measured power consumption was approximately 39 percent above the guarantee. Waste product guarantees are discussed later.

At the guaranteed SO_2 removal the lime stoichiometry was about 20 percent below guarantee, the system pressure drop with 3 SDA's in service was 4 inches WG below guarantee.

The power consumption value presented is the average of hourly values recorded during the day of the tests. The power consumption guarantee is, however, based on a long term average. Periodic measurements of power consumption taken since the performance tests have all been lower and the system is considered to have met the power consumption guarantee.

On the basis of the performance test results and debit/credit calculations, Sunflower and Joy/Niro have proceeded with contract close out.

The baghouses have all operated very well with very few problems. There has been no evidence of bag wetting or blinding due to operation downstream of the spray dryer. The bags have been in service for more than four years at Riverside and have experienced a wide range of test and operating conditions. Antelope Valley has experienced few bag failures during the two years of operation. After two years of operation at Holcomb approximately 2.5 percent of the bags have been replaced.

The successful operation and completion of performance tests for these units supports the Joy/Niro design approach as discussed earlier.

PROBLEM AREAS

The four systems discussed represent the first generation of SDA-FGD systems and as such have experienced controls and equipment problems. The majority of these were corrected during start-up, however, some problems have required more effort to resolve. Generally these problems can be divided into the following: chamber deposits, wheel wear plate cracking, head tank pluggage, limeslaking turndown, conveying systems operation, tank and piping pluggage, and baghouse corrosion. Table 5 presents a summary of problems encountered, corrective actions taken and current status.

TABLE 5 PROBLEMS AREAS AND CORRECTIVE ACTIONS

Component/System	Problem	Corrective Actions	Current Status
SDA chambers	Wall deposits hopper pluggage	- Add feed forward trim to SDA outlet temperature control loop - Adjust gas disperser split ratio - Reduce wheel protection water flow rate - Modify wheel water introduction angle - Install lump breakers at hopper outlet - Improve slurry solids dependability	- Situation greatly improved. - Some continuing wall deposits with system upsets
Atomizer wheel wear plates	Cracking	- Interlock protection water with outlet dampers - Add stress relief bands to wear plates - Spin test all new wear plates - Investigate alternate materials of construction	- Frequency of cracking reduced - New wear plates being monitored
Headtank	Screen plugging	- Increase flow through headtank - Modify screen design	- Problem resolved
Ballmill slakers	Feed chute pluggage at low loads	- Modify feed chute - Increase venting capacity - Operate in on/off mode	- Turndown capacity improved - Some systems operated in of/off mode
Conveying systems	Overloading, wear, flight breakage, plugging	- Clean baghouse on timed sequence - Retension drag chains - Install guide rails - Realign conveyors - Increase conveying air temperature - Replace drag chains with pneumatic systems - Increase conveying capacity	- Overloading corrected - Wear & flight breakage reduced - Some problem remaining with deposits in recycle conveying systems & recycle conveying capacity
Slurry tanks - piping	Solids settling out	- Modify piping design - Modify pump suction design - Lower slurry excess lime by operation at design conditions - Replace pumps - increase pumping	- Piping problems essentially resolved - Tank settling out improved - Some problems remain with tank agitation and plugging
Baghouse	Localized corrosion at access doors at internal corners around duct leaks	- Install new access door gasketing - Ensure tight closing of access doors - Seal duct leaks - Upgrade insulation at cold spots	- Corrosion reduced - Monitoring corrosion

Chamber deposits ranging from localized wall deposits to extensive wall and central gas disperser pluggage have been observed during start-up of the systems. For minor localized deposits, continued operation at design conditions results in a scouring off of the deposits, however, major deposits have, in isolated cases, required shut-down of the SDA for digging out and removal. Causes of these deposits are believed to be due to four main reasons, improper setting of the flue gas flow between the roof and central gas dispersers, slow response of the spray dryer outlet temperature control loop, excessive wheel protection water, and operation at lower than design feed solids concentration.

The flue gas inlet flow split between the roof and central gas disperser is set during start-up. The proper split ratio may range from 60:40 to 50:50 depending on site specific operating parameters. At Antelope and Holcomb it was found that the split was set incorrectly during start-up. Adjustment of the split ratio improved operation and decreased deposit formation.

Initially spray dryer outlet temperatures were controlled based on thermocouples located in the spray dryer outlet duct. It was subsequently found that rapid changes in flue gas flow (i.e. unit load changes from adding or dropping a coal mill) resulted in SDA outlet temperature excursions that may have contributed to deposit formation. A feed forward signal based on inlet gas flow and temperature has been added to the outlet temperature control loop. This has led to more constant SDA outlet temperatures and decreased deposits.

Wheel protection water is added at the top of the atomizer wheel to protect the ceramic wear plates from thermal shock. Initially, the wheel protection water flowrate was set at between 3 and 5 gpm. At these rates some of the water would creep out from the top of the wheel and cause deposit build-up on the skirt above the atomizer. These deposits would periodically break off and fall to the bottom of the chamber resulting in pluggage or fall on the central gas disperser partially plugging the inlet gas flow path. The pluggage would cause a disturbance in the gas flow distribution and result in more deposits being formed. This problem has been resolved by reducing the wheel protection water flowrate to between 0.5 and 0.9 gpm and installing a new type liquid distributor with a modified method of introducing the water. Some systems have had lump breakers installed at the bottom of the SDA cone to ensure that no deposits cause pluggage of the outlet and subsequent solids build-up.

Some deposit formation is attributed to periods of operation at lower than design feed solids concentrations. Reasons for these feed solids excursions are varied, but generally result from an upset in the feed preparation system. For Antelope Valley Unit No. 1, lack of adequate conveying capacity to the recycle system has resulted in periods of low feed solids operation. During these periods the SDA outlet temperature is increased slightly to ensure adequate drying. This problem is being resolved by modifications to the baghouse conveying system.

The combination of these four correction actions has led to greatly improved operation and have essentially eliminated deposit formation.

The atomizer wheels contain upper and lower ceramic wear plates. Several wear plates have cracked and had to be replaced. Initially, the cause of this cracking was believed to be due to thermal shock when cold slurry was introduced to the flue gas heated wear plates. A change in the wheel protection water flow logic was made such that wheel protection water flow begins with the initial opening of the spray dryer outlet dampers. This change substantially reduced the frequency of wear plate cracking;

however, some additional cracking has occurred. Niro Atomizer has since modified the bottom wear plate by shrink fitting a metal ring around the periphery of the wheel to reduce stress. In addition, the manufacturing quality control has been improved and all wear plates are now spin tested at overspeed conditions in a Niro Atomizer test stand prior to shipping. No wear plates that have been spin tested have failed in service.

The atomizer headtanks are equipped with screens for removal of oversized material. These screens have at times partially plugged restricting flow to the atomizers. Corrective actions taken have been to increase the slurry flowrate through the head tank and modify the screen design to increase the slurry velocity. These actions have resolved this problem.

All of the reagent preparation systems include ball mill slakers. These slakers have experienced plugging of the feed chute when operated at low loads. The problem is believed to be due to vapor produced during slaking entering the feed chute and causing the pebble lime to begin to slake and adhear to the chute.

Several modifications have been made to the feed chute and the venting systems. These have improved the operation, but have not completely eliminated plugging problems. To compensate for this the slakers are typically operated in an on-off mode. This mode of operation is considered as satisfactory.

Conveying systems used to transport product at these sites include mechanical drag chain conveyors under the SDA's and either pneumatic or a combination of pneumatic and mechanical conveyors under the baghouse. Conveying system problems have included overloading of mechanical conveyors under the baghouse, excessive wear and paddle flight breakage, lack of capacity of pneumatic conveyors, and scale build-up in the pneumatic conveying lines.

At Holcomb overloading of the conveyors, and subsequent trip-out, was due to excessive solids flow during a sudden loss of flue gas as with an ID fan or boiler trip. The baghouse cleaning mode was changed from pressure drop initiated to time initiated cleaning to smooth out the flow of solids to the conveyor. Control logic has been modified to automatically trip out the baghouse unloading conveyors in the event of an I.D. fan trip. In addition, the baghouse uloading conveyors were slowed down from 30 feet per minute to 20 feet per minute. These changes have eliminated the overloading problem. Resolution of mechanical conveying problems have included retensioning of the drag chain, addition of guide rails on the backside of the conveyors, and realignment of the conveyors. These have resulted in a decrease in the wear and number of broken flights.

At Antelope Valley the all mechanical conveying system for the spray dryers is being modified to a combination of mechanical and pneumatic. The SDA product will be transported to the disposal silo. Platte River Power Authority plans to replace the spray dryer mechanical conveying system at Rawhide with product lump breakers and a dense phase pneumatic transport system to the waste product silo.

The vacuum type pneumatic transport system for the baghouse at Rawhide has suffered from deposit build-ups in elbows of the conveying lines. Chemical analysis of the deposits indicate they consist primarily of lime based compounds. It is thought that the deposits are caused from hot moisture laden gases being drawn into the conveying system along with the solids from the baghouse. Cooling of the gas results in condensation with subsequent wetting of the product. The Rawhide product consists of a reactive flyash which contains unslaked lime and calcium sulfate anhydrite. When dampened these compounds rapidly hydrate to form cementitious materials. Operation at design spray down temperatures will result in a higher recycle ratio and is expected to aid in minimizing this problem. Investigations are continuing to reach a solution of this problem. In addition, the recycle ash conveying system has been modified to increase capacity to ensure adequate solids in the feed.

Problems in the process tanks and piping have been due to solids setting and subsequent plugging. This has been due to lower than design slurry flow rates, dead legs in the slurry piping, and the reactive nature of the flyash especially when high excess lime is present. These problems have resulted in periodic shut-down of the affected components for removal of hardened deposits. Corrective actions have included modification of piping runs to increase slurry velocity and eliminate dead legs, elevating pump suction spools in slurry tanks, modifying agitation and tank baffle design, operation of the SDA system at design conditions, and periodic rotation of equipment operation.

Modifications to the piping system have reduced line plugging substantially. The elevating of pump suction spools and the use of a goose neck arrangement for the suction piping has eliminated pump suction plugging. Operation at design conditions has lowered the excess lime in the slurry and resulted in improved slurry deposit problems in slurry tanks and piping.

All of the systems are designed with redundant pumps, and in some cases piping loops, for slurry service. When the spare systems are allowed to set idle for extended periods of time slurry sometimes leaks into the piping and causes plugging. Periodic rotation of spare equipment into service keeps these systems from plugging. Antelope Valley is replacing slurry cross over connections with single lines to each pump.

The above-mentioned modifications have reduced plugging problems, but further work is still required. Joy/Niro are working together with the utilities and the agitator vendors to complete the resolution of these problems.

Baghouse corrosion is a concern because of the low temperature and high moisture content of the flue gas. The systems in service with the exception of Riverside have shown only minor localized corrosion due to cold spots on baghouse walls. The cold spots are located around access doors, baghouse internal corners, and leaking duct joints. New gasketing material has been installed around the access doors and additional attention is being given to ensuring a tight fit when access doors are closed. Duct leaks are being sealed and insulation is being upgraded at corners and other identified cold areas. It is believed that these fixes will reduce this type of corrosion.

The Riverside baghouse has experienced localized and some general corrosion. As a demonstration facility the baghouse has seen temperatures very close to the dew point, intermittent spray dryer operation, very low gas velocities and low ambient temperatures. The insulation design was for non SDA service where the typical inlet temperatures are close to double those after an SDA. In addition, the boilers are peaking units. As such they change load substantially daily and are banked over the week-ends. Banking the boilers causes baghouse temperatures to pass through the dew point several times each week-end resulting in condensation of acidic water on the walls and subsequent corrosion.

SYSTEM AVAILABILITY

Although the systems have experienced problems, they have maintained very high availabilities. Table 6 presents availability data for the SDA systems.

TABLE 6 SDA AVAILABILITY DATA

	Riverside	**Antelope Valley**	**Holcomb**	**Rawhide**
Availability %	90.4	95	99.5	99
Period	Since Aug. 1984	Since July 1984	Since Jan. 1985	Since March 1984

As can be seen the system avaiabilities have been very high, ranging from 90 to greater than 99 persent. This is very impressive when one considers that these plants represent the first generation of a new technology. These high availabilities are a reflection of the simplicity of the process, the design of the spray dryers and atomizers, the redundancy built into the system, and the dedication of the operating personnel in keeping the systems maintained and on-line.

Riverside availability at 90.4 is in part due to the peaking operation of the boiler which requires many restarts of the system from a cold condition. Due to the reactive nature of the flyash in the feed slurry, the slurry cannot be stored for extended periods, but must be prepared fresh when restarting the system. This results in some problems with feed solids control during start-up. In addition, during this period NSP has incorporated several new equipment items for evaluation for their Sherco III SDA system now under construction.

The Antelope Valley SDA system operating availability has been approximately 95 percent since the June 1984 boiler outage. The availability when one considers meeting compliance has been somewhat lower (approx. 90 %) due to problems with the recycle conveying system. Modifications, as discussed earlier, have been made during the most recent boiler outage, and it is anticipated that the availability will improve even more.

The Holcomb station availability of 99.5 percent is very high, especially, when one considers the reactive nature of the Powder River Basin flyash which enters the system. Since the first of the year the system has been down for only a total of 14 hours. These hours reflect the time necessary to switch from the operating feed system to the spare to avoid setting of up recycle and feed slurries in non-active systems.

The availability of the Rawhide SDA system to operate when called upon has been 99 percent since the beginning of 1985. The SDA system has maintained the boiler in compliance 91 percent of the time the boiler has operated. Problems with the recycle conveying system have caused losses in feed solids and required raising the SDA outlet temperature to avoid deposits. This has resulted in decreases in SO_2 removal and periods of non compliance. Platte River Authority is currently making modifications to the recycle conveying system which should improve the situation.

Overall the availabilities are high and are continuing to improve. There is a good exchange of information between the operating plants' personnel and Joy/Niro. As more experience is gained with SDA FGD system design and operation it can be expected that the system availability will improve even more.

WASTE DISPOSAL

All of the plants dispose of their wastes as a dry or dampened material. In addition Riverside, Antelope Valley and Holcomb stations have active programs for utilization of the waste product. Each plant employs a slightly different method of waste product conditioning and disposal. This section presents a brief description of the waste conditioning and disposal practices at each site as well as a comparison of chemical/physical properties of the products. Properties of laboratory and on-site conditioned product are presented.

The Riverside disposal/utilization practices have been reported previously (9) and will not be discussed here. Waste conditioning is accomplished at Riverside by blending 10 to 15 percent water in either a rotary dustless unloader or a batch type cement mixer. The two systems were installed for evaluation for use at the NSP's Sherburne County Unit 3. Because of the utilization of the waste product only enough water is added to eliminate dusting while loading dump trucks and transporting the product to interim storage sites. From the storage sites the product is transported to utilization sites where additional water is added prior to compaction.

The application of a batch type cement mixer is somewhat unique in the utility industry. The mixer in use at Riverside is manufactured by Mixer Systems, Inc. and has a capacity of 120 tons/hr. Product is introduced from the waste silo via a rotary valve until the mixer is full. The mixer contains paddles which are then rotated. The rotating paddles mix the product with water which is sprayed into the drum. The mixing operation requires 60 seconds and the product discharges from the bottom. Operation to date has been very satisfactory.

The rotary unloader is manufactued by Fuller Co. A powder head tank is installed ahead of the rotary unloader to ensure a uniform flow of product to the unloader. Its operation has also been satisfactory.

Antelope Valley Unit No. 1 disposes of conditioned waste product in mined out portions of the coal mine which supplies the plant. A small fraction of the product is marketed for various utilization projects.

Dry FGD waste is conveyed from one of two 4800 ton capacity storage silos, via air slides, to one of two 225 ton/h pugmill mixers (Dravo-Wellman) where 15-30 percent water is added. Bottom ash from the dewatering bins can also be added to the pugmills. Conditioned product discharged from the pugmills onto conveyors, is conveyed approx. one mile to the initial disposal site, and stacked out in an arc by a portable radial stacker. From there it is spread into lifts and compacted by heavy equipment. The system is operated intermittantly as each storage silo is filled to capacity.

The product from the Holcomb station is disposed of in a land fill located approximately one mile from the boiler. Design of the waste conditioning system and landfill were included in the FGD system scope. The conditioning system at Holcomb consists of two 40 hour capacity waste product silos, a rotary valve and deflection type flow meter for product flow control, two 170 ton/h Dravo-Wellman paddle mixers and a transfer conveyor for truck loading. The conditioned product containing approximately 25 percent added water, is transported in 35 ton ash haul trucks to the landfill where it is spread in two foot lifts and compacted. The landfill which also accomodates bottom ash will ultimately occupy approximately 110 acres at a maximum depth of 50 ft. Minimum unconfined compressive strength and maximum permeability of the conditioned product were included in the FGD system guarantees.

Holcomb station has begun a program to sell their end product. They have a committment for the purchase of 10,000 tons of product in 1985. This product will be used as an admixture to asphalt for use in paving. The amount of product sold is expected to increase in the future.

The waste product from Rawhide is also disposed of in an on-site landfill. Dry product from the waste product silo is transported, via air slides, to closed trucks for transportation to the landfill. There it is spread, via bottom gates on the trucks, dampened, and compacted. A waste conditioning system, very similar in design to the Holcomb system, is available, but suffered from solids build-up during early operation and has been abandoned until the system operation has been stabilized.

A series of samples of endproduct from the four stations were recently collected for evaluation of waste disposal properties. Where possible both dry and conditioned product samples were collected. Table 7 shows the chemical composition of the end products tested.

TABLE 7 ENDPRODUCT COMPOSITIONS

CONSTITUENT %	RIVERSIDE	ANTELOPE	HOLCOMB	RAWHIDE
SiO_2	24	34	28	25
Al_2O_3	10	15	14	16
Fe_2O_3	2.8	5.9	5.7	6.3
CaO	24	18	25	28
MgO	2.8	5.2	4.0	3.1
Na_2O	0.33	5.4	0.9	1.2
K_2O	0.25	0.9	0.3	0.2
SO_3	14	11	12	14
LOI*	21	1.0	5.1	4.3
Moisture	0.5	0.8	0.3	0.4
Bulk density				
Loose g/cm^3	0.62	0.89	0.77	0.71
Tapped g/cm^3	0.73	1.05	0.89	0.89

* Loss on ignition, 815°C

The high loss on ignition (L.O.I.) for the Riverside product is due to the coke blended in the fuel and the age of the boilers. The products all contain a high fraction of calcium and sulfur due to the SDA process. The moisture contents are low indicating proper drying.

Samples of the dry product were returned to the laboratory for determination of optimum moisture and compacted density. Proctor cylinders were prepared, cured at 23°C for 28 days, and tested for unconfined compressive strength and permeability.

Proctor cylinders of conditioned product were prepared on-site and returned to the laboratory for curing and testing. EPA-EP extracts were analyzed for each sample. Table 8 presents the results of these tests.

TABLE 8 ENDPRODUCT DISPOSAL PROPERTIES

	RIVERSIDE	**ANTELOPE**		**HOLCOMB**		**RAWHIDE**
Conditioning	Lab.	Lab.	On-site	Lab.	On-site	Lab.
Moisture %	35	15	17	22	20	29
Density lbs/ft^3	77	113	110	96	93	87
Unconfirmed compressive strength 7 day, psi	1125	650	1270	690	520	770
28 day, psi	1420	1310	2790	1260	760	2020
Permeability 28 day, 10^{-6} cm/sec	0.6	0.023	*	2.6	0.02	0.31
EPA-EP ANALYSIS						
Arsenic mg/l	< 0.002	< 0.001	< 0.001	< 0.001	0.015	< 0.001
Barium	1.2	0.125	0.83	0.083	0.42	0.333
Cadmium	< 0.01	0.038	0.008	0.026	0.019	0.015
Chromium	< 0.06	0.192	0.096	0.192	0.098	0.167
Lead	0.15	1.141	0.211	0.87	0.35	0.422
Mercury	0.007	0.0012	0.0017	0.0012	0.0020	0.0012
Selenium	< 0.002	< 0.001	0.011	0.005	0.020	0.018
Silver	0.01	0.041	< 0.008	0.033	0.023	0.47

* Could not saturate: Approx. 10^{-8}

The results show a wide range of optimum moisture contents as noted for the laboratory samples. The site samples show that the conditioning system closely approach the optimum. The low density for the Riverside product reflect the coke in the fuel.

The unconfined compressive strengths are all high and show continued strength development with time. The difference between the laboratory and site values for Antelope are because the laboratory samples were collected from the baghouse hoppers, whereas, the site conditioned sample was collected from the disposal silo and represents different operating conditions. These values show good agreement with earlier testwork (4).

Permeability coefficient measurements were made using a pressure assisted method. This was due to difficulties in saturating the samples with a standard falling head method. The permeabilities are all low on the order of 10^{-6} cm/sec or less.

Results of the EPA-EP analysis show that all values are below RCRA limits and indicate that all samples would be classified as a non hazardous waste based on toxicity.

In addition to laboratory and site conditioned samples core samples have been taken from the Holcomb landfill site. Table 9 compares results for density, unconfined compressive strength and permeability for these samples.

TABLE 9 HOLCOMB STATION ENDPRODUCT COMPARISON

Type sample		**Laboratory prepared**	**Site conditioned**	**Landfill core sample**
Density	lbs/ft^3	96	93	80
Unconfined compressive strength	psi	1260	760	550
Permeability x 10^{-6} cm/sec		2.6	0.02	3.9

The results show landfill densities of about 85 percent of optimum. This indicates good compaction in the landfill site. The compressive strength is about half of the laboratory values and reflects differences in compaction and curing conditions for the two samples. Compressive strength of the core sample is very high, equivalent to a soil bearing capacity of 40 tons per square foot. The permeability of the core sample although higher than the laboratory value is very low. These results indicate that the landfill is being properly placed and maintained.

CONCLUSIONS

Spray dryer absorption plants supplied by Joy/Niro have successfully started up and are in commercial operation treating flue gases from four coal-fired utility boilers. The start-up and early operation of these four first generation systems have not been without problems common to the application of a new technology. Most problem areas have been resolved and the systems are all operating at high availabilities. Efforts are continuing by the utilities, Joy/Niro and equipment suppliers to successfully resolve the remaining problem areas.

All systems have passed compliance tests. Two plants which have undergone performance guarantee testing have successfully met guarantee requirements. Guaranteed SO_2 removals have been demonstrated at lower than guarantee lime stoichiometries and the overall evaluated operating costs are lower than guaranteed.

The design features of a single centrally located reliable rotary atomizer, a compound gas disperser, two point discharge, product recycle, and operation at high feed solids have been shown to be correct, as indicated by the high availabilities of these four units (90-99 percent).

The endproducts from these systems all exhibit high compressive strengths and low permeabilities. They are being disposed of in landfills or utilized for mine/land reclamation. EPA-EP procedure tests show the product to be a non-hazardous material.

ACKNOWLEDGEMENT

The authors wish to acknowledge the contributions made by the dedicated staff of the utilities involved and Joy/Niro who have worked hard to make these systems successfull.

We would also like to acknowledge American Resources, Inc. for their assistance in conducting the waste disposal investigations.

REFERENCES

1. Janssen, K.E. - Eriksen, R.L.
Basin Electric's Involvement with Dry Flue Gas Desulfurization
EPA FGD Symposium, March 1979, Las Vegas, Nev.

2. Davis, R.A. - Meyler, J.A. - Gude, K.E. - Dry SO_2 Scrubbing
American Power Conference, April 1979, Chicago, Ill.

3. Brown, B. - Fitzpatrick, M. - Sannes, C. - Skinner, M. - Swanson, A.E.
Performance Response of Dry SO_2 Removal System of Riverside Station
American Power Conference, April 1981, Chicago, Ill.

4. Eriksen, R.L.- Antelope Valley Station Dry FGD and Waste Disposal Systems
Riverside Dry FDG Symposium, June 1981, Minneapolis, Minn.

5. Felsvang, K. - Results from Operation of Riverside Dry Scrubber
Riverside Dry FGD Symposium, June 1981, Minneapolis, Minn.

6. Felsvang, K., et al - Patent no. 4,229,249 - Oct. 21, 1980
Process for flue gas desulfurization

7. Donnelly, J.R. - Ellis, R.P. - Webster, W.C. - Dry Flue Gas Desulfurization End-Product Disposal, Riverside Demonstration Facility Experience
EPA/EPRI FGD Symposium, May 1982, Hollywood, Fla.

8. Cox, G.H. - Dry FGD System Design Holcomb Station Unit 1
Joy/Niro Scrubbing Seminar, June 1982, Minneapolis, Minn.

9. Donnelly, J.R. - Disposal and Utilization of Spray Dryer FGD Endproducts
Canadian Electrical Assoc. Seminar on SO_2 Removal, Oct. 1982, Ottawa, Canada

10. McLaughlin, B.R. - Vail, D.L. - Cox, G.H. - Emerson, R.D.
Dry Scrubbing, Recycling Help Conserve Water in Acid Region
1984 Generation Planbook, Power Magazine

11. Eriksen, R.L. - Spencer, H.W. - Donnelly, J.R.
Design and Operating Experience of the Antelope Valley Unit 1 Dry Scrubber
APCA annual meeting, June 1984, San Francisco, Cal.

12. Emerson, R.D. - Dry FGC Systems: Start-up, Performance and Acceptance Tests
Power Engineering, Oct. 84 pg 50-52

13. Blythe, G.M. - Burke, J.M. - Lewis, D.L. - Thompson, C.M.
Field Evaluation of a Utility Dry Scrubbing System
Draft final report - EPRI contract no. RP 1870-4, November 1984

14. Emerson, R.D.
Design and Operating Experience of the Holcomb Station Unit 1 Dry Scrubber
1984 Winter meeting, NUS organization Dec. 84, Clearwater, Florida

15. Emerson, R.D. - Eriksen, R.L. - Operating Experience of Dry FGC Systems
Acid Rain Conference, March 1985, Washington D.C.

PROBLEM SOLVING AND RESULTS
OF PERFORMANCE TEST AT
UNITED POWER ASSOCIATION
STANTON STATION
DRY FGD SYSTEM

BY

Ghassem B. Manavizadeh
Stefan Negrea
(Research-Cottrell, Inc.)

Gordon Westerlind
(United Power Association)

Alan W. Ferguson
(Black & Veatch Engineers-Architects)

ABSTRACT

Use of data acquisition and management techniques of solving problems associated with spray dryer and fabric filter are discussed. Methods of cooperation between Owner, A/E and the System Supplier are highlighted. Results of the first FGD performance test, 60 day reliability test, its inconclusive results and the improvements generated by its data, as well as the preliminary final performance test (one year after the initial test) are presented. A brief description of the system design, field modifications and its operation is presented.

Using methods specified by EPA for compliance testing, guaranteed SO_2 removal efficiency and particulate emission were achieved at the UPA Stanton Dry FGD Unit. Preliminary data are also presented on use of sonic horns for reducing pressure drop in fabric filter and their effect on SO_2 removal and particulate emission.

I. INTRODUCTION

The United Power Association Stanton Plant has a steam generator equipped with a dry flue gas desulfurization (FGD) system. Some difficulties were encountered during startup with the performance of the FGD system. This paper presents a discussion of the cooperative approach used by the involved organizations--United Power Association, Owner; Black & Veatch, Engineer; and Research Cottrell, System Supplier--to resolve the problems.

First, a brief description of the plant and the installed FGD system are described. Other papers (1, 2, 3) are available which provide a detailed description for the reader interested in the details of the equipment design. The next section of the paper presents a discussion of the cooperative approach used to resolve problems encountered. The following section discusses some of the problems and a describes the steps taken to find a satisfactory solution. A summary of the results of the performance testing completed on the FGD system is presented. Also included are discussions of problems encountered and resolved during the performance test period.

II. PLANT DESCRIPTION

The Stanton Station of United Power Association is located near Stanton, North Dakota. The plant burns locally mined lignite and was designed for 185 MW gross electrical output. The steam generator is a natural circulation, front-fired unit. The backpass is divided into parallel reheat and economizer sections. Shortly after being placed in service in 1967, the plant experienced boiler slagging and fouling problems. These problems resulted in the unit being effectively derated to 120 to 130 MW gross electrical output. UPA was unable to continuously sustain loads greater than these.

A supplemental steam generator was added to provide the turbine with sufficient steam to allow operation at the original electrical output. The supplemental steam generator supplies approximately 60 MW of generating capability which restores the plant to its design capacity. The supplemental steam generator is a natural circulation tangentially fired unit which can be operated in parallel with the existing steam generator.

The timing of the supplemental steam generator addition (1979 contract award for steam generator) was such that the EPA ruled that it was subject to the 1979 New-Source Performance Standards Subpart D(a) regulations for particulate, SO_2 removal and NO_x emissions. As a result of this determination, an air quality control system for removal of both particulate and SO_2 was required. A dry FGD system consisting of a spray dryer and a fabric filter was chosen.

The dry FGD system includes a single spray dryer for SO_2 removal. The spray dryer is 38 feet in diameter and about 70 feet tall. Design residence time at full load is about 10 seconds. There is no spray dryer bypass capability. Lime is used as the reagent for SO_2 removal. The spray dryer is equipped with three rotary atomizers. The three rotary atomizers are top mounted and designed for 14,000 rpm. The atomizer wheels are titanium with tungsten carbide inserts in the slurry exit ports. The atomizers are independent and any atomizer can be removed from service while the remaining atomizers are operating. This design feature is important for a single spray dryer system because it does not cause emission exceedences or require shutdowns for atomizer normal inspection or maintenance. In essence, compliance can be maintained at times with two atomizers in service.

The fabric filter is reverse gas cleaned. The cloth velocity (air-to-cloth ratio) is 1.65 ft/min (acfm/ft^2) with all 10 compartments operating. Each compartment contains 300 eight-inch diameter fiberglass bags. The fabric filter is equipped with a full flow bypass for emergency and startup operation.

The design basis coal analyses and operating conditions for the FGD system are shown in Table 1.

The SO_2 removal guarantee varies with fuel composition. A minimum of 70 percent removal or 0.6 LB/MBTU emission must be maintained.

III. TEAM MANAGEMENT APPROACH

United Power Association (UPA) retained Black & Veatch, Engineers-Architects (B&V) to provide design and construction management services for the Stanton Station supplemental steam generator project. B&V was responsible for the engineering and design of the project, and construction coordination of the contractors performing the work in the field. UPA worked closely with B&V during the preliminary and final design phases of the project and also had a full-time site representative assigned to the project during the construction phase.

The contract for furnishing and erecting the dry FGD system was awarded to Research-Cottrell, Inc. (R-C), in January 1980. R-C supplied design information on their equipment and process by participating in design conferences and submitting engineering data and drawings to B&V. R-C also procured the equipment and provided field erection and startup services, including testing required to place the dry FGD system in operation. The project schedule called for a May 1981 construction start date with the system being ready for commercial operation by September 1982. The design and construction phases of the dry FGD system proceeded as scheduled. Although these phases of the project were not trouble free, the design and erection related problems were addressed in a timely manner and project schedule was maintained. Construction of the dry FGD system started in May 1981 and construction completion allowed startup and testing to begin during July 1982.

TABLE 1

FGD DESIGN BASIS COAL ANALYSIS

Heating Value (BTU/LB)	5976-7367
Moisture (percent)	27.9-37.9
Ash (percent)	4.9-11.8
Sulfur (percent)	0.33-1.94
Volatile Matter (percent)	25.11-31.96
Fixed Carbon (percent)	29.43-34.42

FGD DESIGN CONDITIONS

	Typical 0.77% Sulfur	Maximum 1.94% Sulfur
Gas Flow (ACFM)	284,184	331,567
Nominal Gas Inlet Temperature (^{o}F)	323	323
SO_2 Inlet Concentration (PPM)	788	1,988
Particulate Inlet Flow Rate (LB/H)	5,060	10,140
SO_2 Removal Efficiency (Percent)	72.7	90.5
Particulate Emission Rate (LB/MBTU)	0.03	0.03
SO_2 Emission Rate (LB/MBTU)	0.6	0.6

The contractual relationships and responsibilities established between UPA, B&V, and R-C on this project were not unique; many projects employ similar relationships between the Owner, Engineer, and System Supplier. Although the contractual arrangement between parties participating in the project is not unique, the high degree of cooperation between UPA, B&V, and R-C, especially during the startup phase of the project, was quite significant to the success of the project. Problems with the dry FGD system were experienced during startup. However, rather than taking the position that correction of system deficiencies was entirely the responsibility of R-C, B&V and UPA placed importance on resolving the system deficiencies as quickly as possible and initiating a team approach be used for problem resolution.

The approach used to resolve the dry FGD system problems involved active participation by UPA, B&V, and R-C. As problems with the system were identified, they were discussed and corrective was proposed and action agreed upon as soon as possible after a problem was detected. Meetings, both on site and at UPA's corporate headquarters were held to review the status of system corrections and discuss newly encountered problems. The number of attendees at these meetings was large at times, but all firms wanted personnel attending the meetings that could contribute to resolution of the problems. The meetings were extremely beneficial in allowing UPA's plant operating staff, which was operating the dry FGD system under R-C's direction, to explain the problems being encountered and to offer their recommendations for system improvements from an operator's viewpoint. R-C used the meetings to discuss their recommendedcorrective action for the system problems and to obtain authorizationfrom UPA and B&V to proceed. The meetings were not held on a regularly scheduled basis, but rather were convened as required to address the problems as they were encountered.

Another example of cooperation among the firms in resolving system problems was UPA's willingness to make the dry FGD system modifications proposed by R-C, especially those involving the spray dryer, requiring outages of the dry FGD system to implement. Dry FGD system outages were granted whenever possible, even though outages required the supplemental steam generator to be shut down because of the lack of capability to bypass the spray dryer. Modifications to the dry FGD system such as atomizer modifications, that did not require shutdown of the supplemental steam generator but resulted in emission violations required UPA to coordinate these activities with the North Dakota State Department of Health (NDSDH). UPA's permit from the NDSDH required that emissions from the supplemental steam generator be in compliance with applicable regulations within 180 days after initial coal firing began. Modifications to the dry FGD system continued past this 180-day period, which required UPA to work closely with the NDSDH whenever emission limits were being exceeded. Even though many of the early modifications proposed by R-C were experimental and were notentirely successful, having the dry FGD system available to try innovative solutions and test proposed modifications greatly reduced the time required to arrive at the final system modifications.

Progress of the dry FGD system corrections was tracked through a weekly status report that was prepared and distributed by R-C. The reports provided information on the status of all deficiencies previously identified and also listed new problems encountered since the last project status meeting and the proposed corrective action.

Resolution of system problems and reliable operation of the dry FGD system was given very high priority by all firms involved. Senior management personnel from UPA, B&V and R-C were actively involved in all phases of problem resolution. This high level involvement allowed decisions and commitments to be made on the spot and was instrumental in cutting through the red tape that normally exists when major expenditures are being made for modifications and rework.

In order to closely monitor performance of the dry FGD system and to evaluate various system parameters, R-C has installed a remote control monitoring system at their office in New Jersey. This facility allows online monitoring of all primary control and performance parameters through duplication of the Modicon 584 control system. This technique was valuable in permitting a continuous understanding of the plant performance and helping to identify problems.

In summary, no new or magical techniques were used to resolve the problem with the dry FGD system. The problems were, however, given high priority and addressed in a logical, methodical manner by the right people in each organization. The skills and resources of UPA, B&V, and R-C were employed as required to meet a common goal--reliable operation of the dry FGD system.

IV. MAJOR PROBLEM AREAS

It is the intent of this paper to discuss the major problem areas associated with the spray dryer/fabric filter, the source of the problems and the solution to those problems.

Fabric Filter

The startup procedure developed by R-C called for the stable operation of the fabric filter followed by startup of the spray dryer. The intention was to coat the bags with fly ash before exposing them to FGD solids, reduce the spray dryer temperature gradually through injecting water and then lime slurry, therefore minimizing the ash loading on the fabric filter. The fabric filter was put into operation in July 1982, without the spray dryer in operation. The fabric filter was designed to operate with the spray dryer in operation. Without spray dryer operation, the fabric filter inlet volume exceeded the design volume by about 25 percent; therefore, the pressure drop across the fabric filter exceeded the design values and limited boiler output. In addition, the reverse gas fan capacity at this condition was marginal. Reverse gas volume dropped below the minimum required value to clean the bags at 8 to 10 inches of pressure drop, and since bags would not clean, the pressure drop could not be controlled and increased unchecked.

Several steps were taken to corrrect the fabric filter startup problems. First, the reverse gas fan wheel size was increased to allow higher static pressure capability of the fan. Second, two sonic horns were placed in each compartment in August 1982. The bigger wheel allowed correct cleaning and reduced the fabric filter pressure drop to below the design value without the spray dryer in operation. Finally, in January 1983, with the successful operation of the spray dryer, the gas volume to the fabric filter was reduced to its normal design volume and thus the fabric filter pressure drop was brought well within the anticipated levels.

Another major problem associated with the fabric filter was reverse gas fan condensation. The severe winter weather in North Dakota causes condensation of the flue gas which leaks through the dampers of the idle reverse gas fan. Research-Cottrell modified the reverse gas fan cleaning operation so that at 130^{o}F and below both fans are running. The other weather-related problem with the fabric filter was the bag damage from condensation. Cold air infiltrates under insulation at the doors. Some cold spots were as low as 120^{o}F. To resolve this problem, 3 foot high band of insulation material was added around the fabric filter exterior at the cell plate level, and additional insulation was added around the door cavity. Although condensation has been greatly reduced by these solutions, occasional incidents could exist if the doors are not properly closed.

Liquid-Gas Distribution

First operation of the spray dryer was during September 1982. It was soon observed that spray dryer wall temperature varied from side-to-side, even at a constant outlet temperature. Additionally, operation of atomizers with lime slurry caused a substantial (12 to 24 inches) wall buildup. Lime slurry was reaching the spray dryer wall before it had the chance to dry. This was not totally unexpected since similar buildup on walls (at times requiring blasting) have been reported by others. There were four potential causes for the wall wetting and uneven temperature distributions:

1. Uneven gas split to the three atomizers.
2. Uneven lime slurry distribution to the three atomizers.
3. Uneven gas distribution around each atomizer.
4. Improper liquid and gas mixing.

Combinations of the above caused zones of high and low gas flow around the wheel. Uneven slurry flow to each of the high and low gas flow zones was contributing to uneven wall temperature distributions and wall buildup.

As problem areas were identified, an intensive program was undertaken to identify acceptable solutions. A key element in solving the above problems was the gas dynamics model testing of the spray dryer. From the model testing, it was soon discovered that the gas inlet plenum of the as-built dryer was not distributing the gas evenly to each atomizer. More flow was thus going toward one atomizer than toward the other two and gas was flowing down the far side of each atomizer, rather than being evenly distributed around each machine. The profile of the vane rings further distorted the gas flow. The model was modified to simulate the as-built unit and calibrated against the field before potential modifications were tested.

Testing was done on five different configurations. Each was assessed for gas distribution and pressure drop over the full range of boiler load. As the result of modeling, four modifications were recommended and implemented:

1. Installing two partition walls in the inlet plenum to split the gas flow to three individual ducts (Figure 1). This corrected the uneven gas split to each atomizer.

2. Turning vanes were added above each vane ring to evenly direct the gas down (Figure 2). This corrected the uneven gas distribution around each atomizer.

3. Lowering the machines.

4. Adding skirts around the outer circumference of the vane ring. The last two modifications brought the liquid and gas closer and further improved the distribution around the vane ring.

At the time these modifications were implemented, the spray dryer walls were jack-hammer cleaned in December 1982,. Since then, temperature gradient has been good and no wall buildup has been observed.

Reagent/Water

The Stanton Station dry FGD system was designed to use lime slurry as a reagent to remove sulfur dioxide from the flue gas. Lime from the lime silo is fed to a paste slaker to produce 20 percent solids lime slurry. Lime slurry is pumped to the spray dryer where it mixes with dilution water for temperature control prior to atomization. Stanton Station uses Missouri River water, which has a hardness of about 250 ppm. The hardness of the water reacts

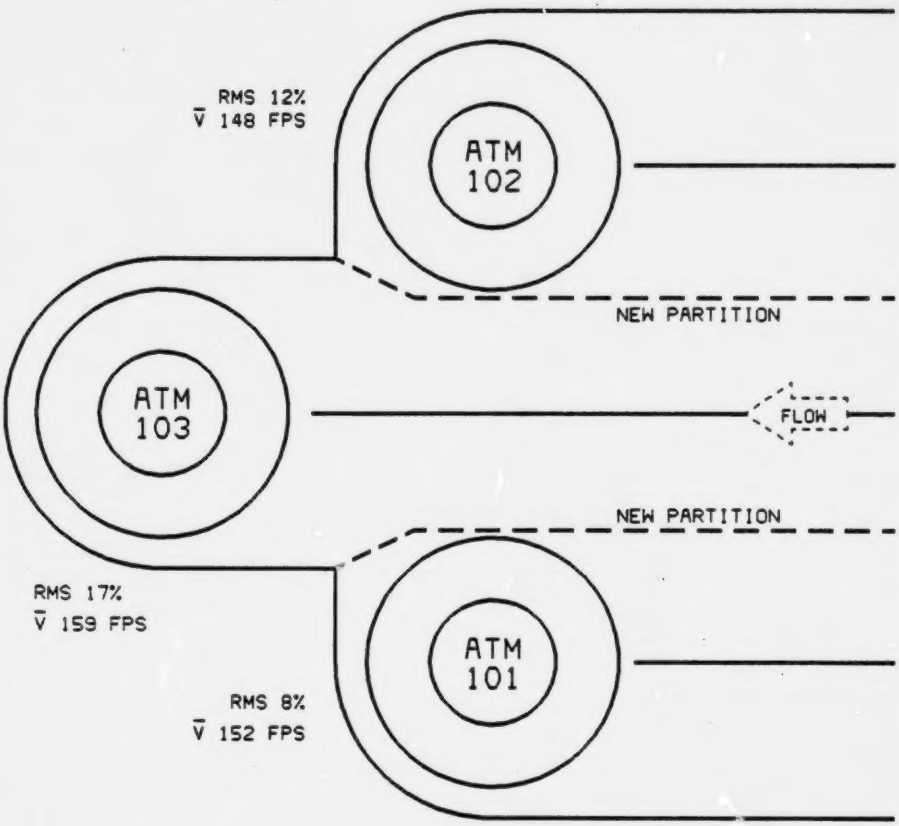

RMS 12%
$\bar{V}$ 148 FPS
ATM 102
NEW PARTITION
ATM 103
FLOW
NEW PARTITION
RMS 17%
$\bar{V}$ 159 FPS
ATM 101
RMS 8%
$\bar{V}$ 152 FPS

FIGURE 1

DISCHARGE FLOW UNIFORMITY

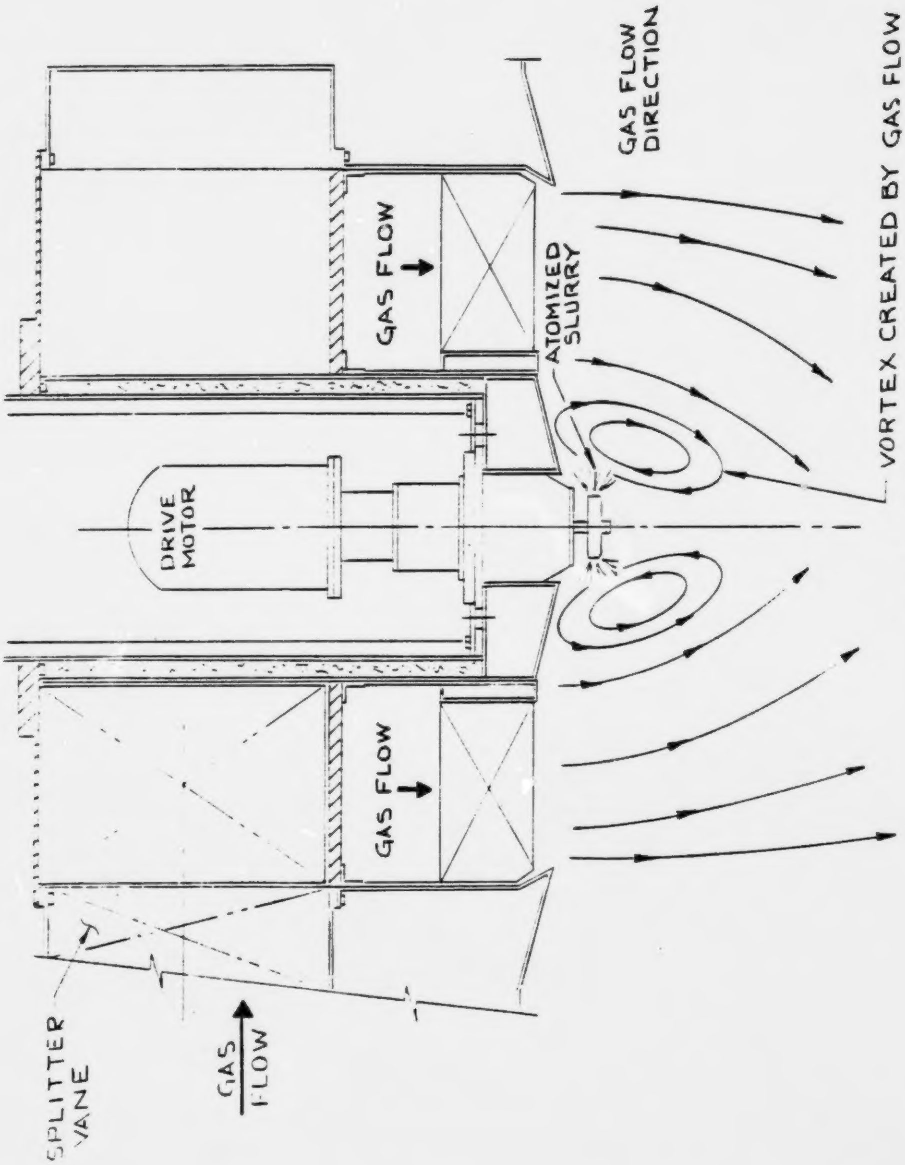

SPLITTER VANE
GAS FLOW
GAS FLOW
GAS FLOW
DRIVE MOTOR
ATOMIZED SLURRY
GAS FLOW DIRECTION
VORTEX CREATED BY GAS FLOW

FIGURE 2

IMPROPER LIQUID AND GAS MIXING

with $Ca(OH)_2$ in the lime slurry in the pipe and produces calcium carbonate, which sticks to the inside of the piping. The buildup inside the pipe accumulated to such an extent that the slurry flow to the atomizers became restricted.

Initially, acid was used to unplug the lines. However, acid flushing started to wear out the lime coating inside the atomizer wheels, which caused atomizer imbalance and forced the atomizer to shut down.

The permanent solution to this problem was to soften the water prior to injection into the piping. The water treatment system at the UPA Stanton Station consists of an enclosed tank to which a small amount of lime slurry is added to the Missouri River water, where it reacts with the water hardness.

The major problem with the reagent was lime contamination. Occasionally, foreign objects in the lime such as wheat, rubber, and stones were carried to the atomizer wheel and caused unbalancing of the atomizer. These foreign materials were believed to be carried from the truck that delivered the lime. Once the trucking company was changed, these problems disappeared.

Wheel/Atomizer

The main problem associated with the atomizers was buildup on top of the atomizer wheel. This buildup caused several machine failures.

Redesigning of the feed slots and cone minimized buildup without the rubbing on the machine (Figure 3). The shroud was also modified to blow the motor cooling air (500 CFM) on top of the wheel, which minimized the solids build-up. Unbalanced liquid flow into the atomizer machine also caused vibration and bearing/shaft failure. Bentley Nevada vibration monitoring equipment was added to warn of unbalance that indicated plugging and wear of the atomizer wheel.

Spray Dryer Interior

Two main problems were associated with the spray dryer interior. The first problem was the solids buildup on top of the outlet duct. The top of the outlet duct located halfway down the spray dryer is exposed to moist solids. These moist solids accumulated on top of the outlet duct and produced a stalagmite shape which eventually reached 8-10 feet in height.

The first attempt to solve this problem, was to place a steep "tent" shape on top of the round outlet duct. This solution proved to be ineffective. The next attempt was to provide an air cannon which blow through a pipe on top of the duct. This fix was quite effective and eliminated all buildup.

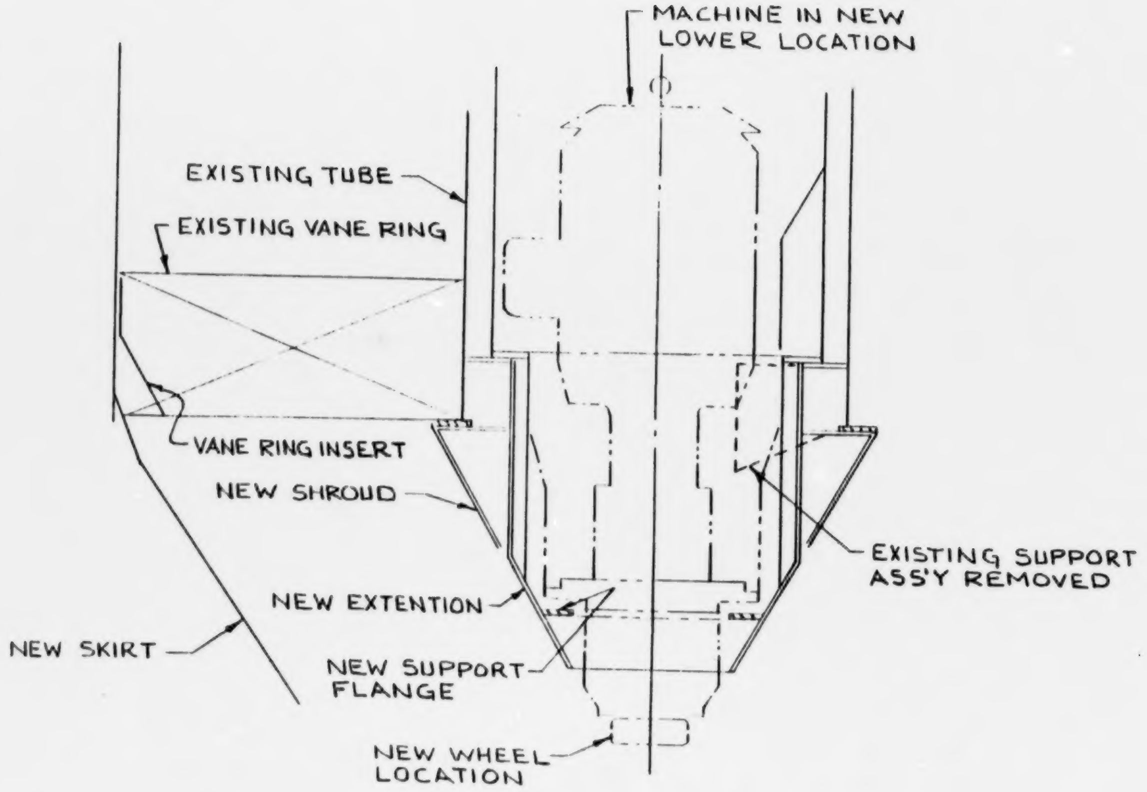

MACHINE IN NEW LOWER LOCATION
EXISTING TUBE
EXISTING VANE RING
VANE RING INSERT
NEW SHROUD
EXISTING SUPPORT ASS'Y REMOVED
NEW EXTENTION
NEW SKIRT
NEW SUPPORT FLANGE
NEW WHEEL LOCATION

FIGURE 3

NEW ATOMIZER LOCATION AND NEW SKIRT

The second problem associated with the spray dryer was that pieces of solids would fall from walls and clog the 4-inch discharge valve. A delumper was placed on the top of the discharge valve and is operates intermittently during the ash removal.

V. PERFORMANCE AND TESTING

A. Reliability Test

After all the system modifications, in December 1983 through January 1984, the dry FGD system successfully passed the 60 day equipment reliability test. During the reliability test, all major equipment operated within full range of boiler load without any failures and emissions were maintained in compliance.

B. Performance Tests

1. Pre-testing: In August 1983, R-C contracted Clean Air Engineering, Inc., to conduct pre-performance test. The objective of this test was to fine tune the dry FGD system prior to the actual performance test. The results of this test are presented in Table 2. They indicate that the dry FGD system is capable of achieving all the guarantee values over the range of boiler design conditions. Conclusions from the pre-testing indicate that no modification is necessary to achieve the required SO_2 removal efficiency with SO_2 emission rate, or particulate emission rate. Lime consumption and utilization were reported to be well within the guarantee values.

2. Performance Test #1: In October 1983, performance tests were conducted to determine if the dry FGD system at Stanton Station was functioning in accordance with the contract guarantees.

The System Performance Test Program included a determination of the following performance parameters:

- Sulfur dioxide removal efficiency.
- Sulfur dioxide emission rate.
- Particulate emission rate.
- Pressure losses.
- Lime Consumption.
- Lime utilization.

TABLE 2

Pre-Testing Summary

August 1983

Parameter	50 Percent Rated Capacity		100 Percent Rated Capacity		Maximum Capacity	
	Guarantee[a]	Test Results	Guarantee[a]	Test Results	Guarantee[a]	Test Results
Total Flue Gas Flow Rate, Actual Wet CFM	136,940	182,050	263,120	271,650	285,880	291,110
SO_2 Removal Efficiency, Percent	70	89.0	70	80.6	70	80.5
SO_2 Emission Rate, LB/MBTU	0.60	0.13	0.6	0.29	0.6	0.35
Particulate Emission Rate, LB/MBTU	0.03	0.029	0.03	0.014	0.03	0.004
System Pressure Losses, in H_2O	6.6		9.3		9.8	
Lime Consumption, LB/MBTU	1,325		1612		1622	
Lime Utilization, Moles SO_2 Removed per 100 Moles CaO Consumed	36		54		58	

(a) Guarantee values are for typical fuel.

Flue gas testing was performed by Energy and Environmental Measurement Corporation (EEMC). Fuel and waste analyses were performed by Minnesota Valley Testing Laboratories, Inc. Steam generator and air quality control system operating data were recorded by UPA and B&V.

The following test methods were agreed to by all parties for determining compliance with the contract guarantees.

MEASUREMENT	METHOD
SO_2 Concentration	Absorber inlet--EPA Method 6 ID fan outlet--EPA Method 6
Particulate Concentration	Absorber inlet--EPA Method 17 ID fan outlet--EPA Method 5
Pressure Loss	Absorber inlet--Kiel Probe measurement and velocity traverse Fabric Filter outlet--Kiel Probe measurement
Lime Consumption	Slurry tank level and slurry specific gravity

The results of these performances, shown in Table 3, indicated at the following contract guarantees were not met at that time.

a. So_2 removal efficiency at 100 percent rated capacity (67.5 percent vs. 70 percent)

b. Pressure loss at 100 percent and maximum capacity

c. Lime consumption and lime utilization at 100 percent and maximum capacity

The SO_2 removal efficiency, though under 70 percent was not considered a major problem since the unit was capable of removing higher than the guarantee value at maximum load. The SO_2 removal efficiency was believed to be caused by instrument error (SO_2 analyzer was set at exactly 70 percent; it should have been set at a higher level to compensate for the analyzer fluctuation).

Higher lime consumption and lime utilization were attributed to inferior lime quality. Lime reactivity examination showed that the lime used during the performance testing was below the specified lime reactivity (55 percent CaO vs. 95 percent).

As a result of higher pressure losses in the system, R-C installed two sonic horns in each of the fabric filter compartments. A new performance test was rescheduled.

TABLE 3

Performance Test #1 Summary

October 1983

	50 Percent Rated Capacity		100 Percent Rated Capacity		Maximum Capacity	
Parameter	Guarantee [a]	Test Results	Guarantee [a]	Test Results	Guarantee [a]	Test Results
Total Flue Gas Flow Rate, Actual Net CFM	136,940	183,850	263,120	252,350	285,880	294,350
SO_2 Removal Efficiency, Percent	70	81	70	67.5	70	71
SO_2 Emission Rate, LB.MBTU	0.60	0.25	0.6	0.47	0.6	0.44
Particulate Emission Rate, LB/MBTU	0.03	0.027	0.03	0.025	0.03	0.017
System Pressure Losses, in H_2O	6.6	-	9.3	9.98[b]	9.8	10.77[b]
Lime Consumption, LB/HR	1,325	951 (1,314)[c]	1612	1,122 (1,969)[c]	1622	1609 (2416)[c]
Lime Utilization, Moles SO_2 Removal per 100 Moles CaO Consumed	36	37	54	45	58	38

(a) Guarantee values are for typical fuel
(b) System pressure loss measurements have been adjusted to design gas flow rates
(c) Lime usage rate adjusted to design inlet SO_2 rate at tested utilization

3. Performance Test #2. Final performance tests were conducted in May 1984. They show that all the guarantees are met at 50 percent and 100 percent boiler load. Tests were also conducted at the maximum boiler load although the gas flue temperature was significantly higher than specified (402°F versus 338°F). Under these conditions, all performance guarantees were met except lime consumption, which could not be demonstrated because of higher spray dryer outlet temperature which lowers lime utilization. Performance test results are summarized in Table 4.

It was judged that the system performed within the guarantee values at the design conditions. The dry FGD system was judged to have satisfied all contractural guarantees for the first test series. The second and final test series is scheduled to be performed in May 1985.

An advantage of the multiple atomizer concept is the degree of redundancy it provides. Operation at compliance SO_2 emission levels with only two atomizers in service was anticipated before startup. In fact, two atomizer operations have been evaluated and have demonstrated guarantee levels of SO_2 removal at full boiler load.

It is important for continuous operation of the dry FGD system that a truly dry product be produced. Dryness is critical to good operation of both the ash handling system and the fabric filter. Since the December 1982 modifications, dry product has been consistently produced. Typical moisture values are 4 percent or less. The pneumatic ash removal system has operated trouble-free since installation of a solids delumper on the spray dryer hopper. The fabric filter has oprated without incident since the introduction of FGD solids. The dryness of the solids is also evident by the absence of fabric filter hopper plugging.

VI. SUMMARY

The first utility-operated, lime reagent, spray dryer/fabric filter system is in operation. During the first year of its operation, problems were encountered with uneven gas/liquid distribution, improper liquid/gas mixing, higher pressure drop in fabric filter, reagent contamination, pipe plating, buildup on the atomizer wheel, solids buildup on the outlet duct, and frequent atomizer vibration.

The approach used to resolve these problems as well as the solutions to these problems were highlighted. Within one year after the initial startup, most of the problems associated with the dry FGD system were resolved. The conclusions are as follows:

TABLE 4

Performance Test #2 Summary

May 1984

Parameter	50 Percent Rated Capacity		100 Percent Rated Capacity		Maximum Capacity	
	Guarantee [a]	Test Results	Guarantee [a]	Test Results	Guarantee[a]	Test Results
Total Flue Gas Flow Rate, Actual Wet CFM	136,940	154,340	263,120	244,900	285,880	286,000
SO_2 Removal Efficiency, Percent	70	80.5	70	74	70	78
SO_2 Emission Rate, LB/MBTU	0.6	0.35	0.6	0.48	0.6	0.48
Particulate Emission Rate, LB/MBTU	0.03	0.02	0.03	0.028	0.03	0.026
System Pressure Losses in H_2O	6.6	5.0[b]	9.3	7.5[b]	9.8	9.6[b]
Lime Consumption LB/MBTU	1,325 (1,316)[c]	1,109	1612	1,367 (1,595)[c]	1622	2,183 (2,037)[c]
Lime Utilization, Moles SO_2 Removed per 100 Moles CaO consumed	36	39	54	54	58	47

(a) Guarantee values are for typical fuel
(b) System pressure loss measurements have been adjusted to design gas flow rates
(c) Lime usage rate adjusted to design inlet SO_2 rate at tested utilization

1. The Stanton Station dry FGD system consistently achieves the required 70 percent SO_2 removal at the specified lime consumption and lime utilization.

2. The sonic horns effectively enhance the bag cleaning in the fabric filter without adverse affect on the fabric filter SO_2 removal.

3. The dry FGD system successfully passed the 60 day equipment reliability test.

4. Performance tests show that all the guarantees, SO_2 and particulate emissions, lime consumption, and pressure drops are met at 50 and 100 percent as well as at MCR boiler loads. Lime consumption was not demonstrated at MCR boiler load due to high spray dry outlet temperature.

5. An advantage of the multiple atomizer concept is the degree of redundancy it provides. Operation at compliance SO_2 emission level can be maintained with one atomizer out of service.

ACKNOWLEDGEMENTS

The authors wish to thank Mr. D. C. Kettner and Mr. D. Taylor of United Power Association and Mr. L. Pflug and Mr. M. Partow of Research-Cottrell, Inc. and Mr. R. W. Dutton of Black & Veatch for their assistance and cooperation during the optimization of the dry FGD system.

REFERNCES

1. D. Taylor, R. W. Dutton and P. H. Dhargalkar, "Performance of a Multiple Atomizer Dry Flue Gas Desulfurization System at United Power Association's Stanton Station," presented at the American Power Conference, Chicago, Illinois, April 1985.

2. R. Roop and L. Pflug, "Dry FGD at United Power Association Stanton Station," presented at the Air Pollution Control Association Conference, San Francisco, California, June 1984.

3. R. W. Dutton and D. C. Kettner, "Supplemental Steam Generator for UPA Stanton Generation Station," presented at the American Power Conference, Chicago, Illinois, April 1983.

HIGH-SULFUR SPRAY DRYER EVALUATIONS

by:

Russell F. Robards and
Robert W. Aldred
Tennessee Valley Authority
1160 Chestnut Street Tower II
Chattanooga, Tennessee 37401

Thomas A. Burnett and
Lynn R. Humphries
Tennessee Valley Authority
501 Chemical Engineering Building
Muscle Shoals, Alabama 35630

Michael J. Widico
Research-Cottrell, Inc.
P.O. Box 1500
Somerville, New Jersey 08876

ABSTRACT

The Tennessee Valley Authority (TVA) initiated a program in 1983 to evaluate a pilot lime spray dryer/baghouse flue gas desulfurization system treating flue gas resulting from the combustion of high-sulfur coal (3.2% S as-fired) at the Shawnee Test Facility. Testing was performed to determine the effects of the major spray dryer operating variables on the spray dryer, the baghouse, and the overall SO_2 removal efficiencies of the system. Variables which were evaluated include lime stoichiometry, approach-to-saturation temperature, recycled solids, flue gas residence time in the spray dryer, atomizer speed, and inlet flue gas temperature. As expected, the SO_2 removal efficiency is a strong function of lime stoichiometry and approach-to-saturation temperature. Recycled solids and the inlet flue gas temperature are also important while atomizer speed and residence time seem to have little, if any, impact over the ranges tested. The most important conclusion is that a spray dryer/baghouse flue gas desulfurization system can attain a sufficiently high SO_2 removal efficiency to meet a 1.2 lb SO_2/MM Btu emission regulation (currently imposed at TVA's Shawnee Steam Plant) when treating flue gas resulting from the combustion of high-sulfur coal. However, careful consideration of the major independent variables is necessary to attain this goal.

INTRODUCTION

TVA's Shawnee Test Facility (STF) Program has provided considerable information on flue gas desulfurization (FGD) systems to the utility industry during the past 14 years of operation. During most of this period, efforts have been directed toward the development of wet FGD systems for eastern high-sulfur bituminous coal applications. With wet FGD reaching commercialization, TVA has redirected its attention to potentially more economical methods to meet future SO_2 compliance regulations, particularly those which may come about as a result of the acid rain controversy. One of the most promising of these new FGD technologies is the dry flue gas desulfurization (DFGD) process (using lime in a spray dryer).

Various utilities, FGD system vendors, Federal agencies, and the Electric Power Research Institute have evaluated spray drying technology on flue gas resulting from the combustion of low-sulfur coal. The results of this previous work have been used in the design, construction, and commercial operation of several full-scale, low-sulfur coal, utility applications. Although widely applied for low-sulfur cases, several concerns have been expressed about the applicability of this technology to high-sulfur coal and thus only a very limited amount of data has been generated with high-sulfur coal. Because of the potential for significant cost saving on the TVA system, TVA has independently decided to evaluate DFGD technology in a pilot plant installed at the Shawnee Steam Plant.

In 1982, negotiations were undertaken between TVA and Research-Cottrell (R-C) to test an E-beam system (for NO_x removal) in conjunction with a spray dryer/baghouse system. As part of the resulting contract, TVA could utilize the spray dryer/baghouse combination for its internal DFGD evaluation. Erection of the equipment at Shawnee was completed in the spring of 1983 and testing has continued during the past two years.

The major project objectives for the spray dryer/baghouse evaluation have been:

- To define the applicability of DFGD technology to treat flue gas generated from burning high-sulfur coal.
- To determine the baghouse contribution to SO_2 removal.
- To perform tests to determine the applicability of spray drying technology to TVA plants under a variety of operating conditions.

During the two years of testing, much data have been generated; too much to be presented here. Discussions will therefore be limited to the basic evaluation of the major DFGD variables. Further dissemination of results will be made at a later date.

The results to be presented here deal with the effects of the main process variables on the SO_2 removal efficiency in the spray dryer absorber (SDA), the baghouse, and the overall system. These variables include lime stoichiometry, approach-to-saturation temperature, inlet flue gas temperature, spray dryer residence time, atomizer speed, and quantity of recycled solids. Also discussed is how these results can be used in the design of a full-scale DFGD system for a high-sulfur coal application.

OVERVIEW OF SYSTEM

The R-C DFGD facility (shown schematically in Figure 1) was installed at TVA's Shawnee Steam Plant during late 1982 and early 1983. It was erected at the STF site adjacent to the existing wet scrubber trains with the capability of receiving flue gas from either unit 9 or 10. Unit 10 typically burns a western Kentucky bituminous coal containing 3.5 to 4.0 percent sulfur (on a dry basis) and the resulting flue gas to the pilot plant contains 2,500 to 3,000 ppm SO_2.

The flue gas from the boiler passes through an extended length of duct from the powerhouse wall out to the R-C facility. The flue gas enters an indirect steam reheater to reheat the flue gas to normal operating temperatures (i.e., to compensate for heat losses in the ductwork) or as specified in the test plan conditions. From the reheater, the flue gas enters the SDA through an inlet scroll located at the top of the SDA. The SDA is a Komline-Sanderson design, is 8 feet in diameter, and has a straight-side height of 13 feet. At the normal gas residence time of 10 seconds, this SDA has a design inlet flue gas rate of 5,000 aft^3/min. The combination of the inlet scroll and the turning vanes installed in the scroll impart a swirl to the flue gas as it enters the SDA through a concentric ring around the rotary atomizer.

An alkaline feed slurry composed of both freshly slaked lime and recycled FGD waste is introduced into the SDA through the rotary atomizer. The slurry flows down through the atomizer and enters the rapidly rotating (approximately 14,000 rpm) wheel. The slurry is forced through the hollow silicon carbide-lined inserts to the outside of the spinning wheel and is

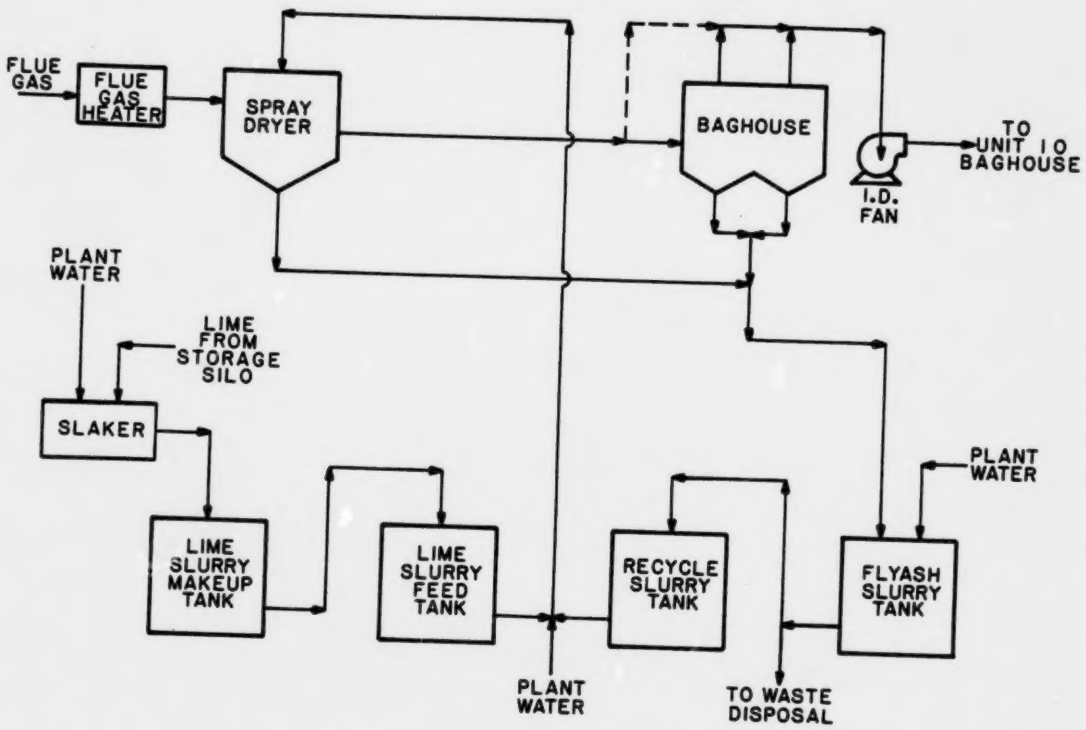

FLUE GAS
FLUE GAS HEATER
SPRAY DRYER
BAGHOUSE
TO UNIT 10 BAGHOUSE
I.D. FAN
PLANT WATER
LIME FROM STORAGE SILO
SLAKER
LIME SLURRY MAKEUP TANK
LIME SLURRY FEED TANK
RECYCLE SLURRY TANK
FLYASH SLURRY TANK
PLANT WATER
PLANT WATER
TO WASTE DISPOSAL

FIGURE 1. R-C FACILITY FLOW DIAGRAM

sheared off as a mist of fine (approximately 50-100 microns) droplets. This intimate mixing of the fine alkaline slurry droplets and the hot, SO_2-laden flue gas results in both the reaction and removal of SO_2 and the evaporation of the water in the slurry. As a result, when the flue gas leaves the SDA through the side exit ductwork, the water in the feed slurry has been evaporated and most of the slurry solids and the fly ash leave the SDA as "dry" particulate matter entrained in the flue gas. Because of the swirl imparted to the flue gas as it enters the SDA and the conical hopper at the base of the spray dryer, the SDA acts as a low efficiency mechanical separator. Most of the larger particles of DFGD waste and fly ash drop out in this conical hopper base of the SDA, enters a rotary air lock, and then drops to an enclosed belt conveyor which transports it to a waste reslurrying tank.

The flue gas from the SDA passes to a 2-compartment, reverse-air, utility-type baghouse. This baghouse initially had 16 full-size utility bags (commercial Teflon-coated fiberglass--1 foot in diameter and 30 feet long) in each compartment and an air-to-cloth ratio of 1.37 $aft^3/min/ft^2$ of cloth area with both compartments online. Since the Shawnee full-scale baghouses were designed for a higher air-to-cloth ratio (approximately 2.0 aft^3/min/ft^2 of cloth area), four bags were removed from each of the pilot-plant baghouse compartments prior to the start of the test program to increase the air-to-cloth ratio to approximately 2.0 and thus more closely simulate the full-scale baghouses.

The entrained particulate matter from the SDA is removed in the baghouse. The flue gas then passes through the ID fan and enters the return ductwork to reenter the unit 9 or unit 10 flue work at the powerhouse wall. The flue gas then passes through the full-scale baghouses and enters the stack breeching and exits through the west stack.

The waste material which collects on the bags is periodically removed during the baghouse reverse-air cleaning cycle. It should be noted that the baghouse cleaning cycle is not typical of a full-scale utility, reverse-air baghouse in that the baghouse is bypassed during the cleaning cycle and both compartments are cleaned sequentially while the baghouse is being bypassed. The waste falls from both compartments into one of two hoppers (one per compartment) at the base of the baghouse. Each of these hoppers is connected to a common screw conveyor. The waste is moved from these hoppers to the waste reslurrying tank by the screw conveyor. As necessary for use in the process, the waste slurry from the reslurrying tank is pumped to the recycle feed tank from which it is pumped to the atomizer. When not needed in the process (i.e., when the recycle feed tank is full), this material is pumped through existing waste disposal lines to the active ash pond for ultimate disposal with the fly ash and bottom ash from the boilers.

Fresh pebble lime is received by truck and pneumatically conveyed to a storage silo. Pebble lime is removed from the silo and fed to the paste-type slaker as needed to prepare a 25 percent solids lime slurry. The degritted slurry is pumped to a 4,000-gallon storage tank where it is stored. When the lime feed tank at the spray dryer is nearly empty, fresh lime slurry is pumped over to refill the lime feed tank. Because this slaking system was originally designed for a 10-MW equivalent wet scrubber, it is overdesigned for feeding the spray dryer system and thus, is operated in a batch mode.

VARIABLES EVALUATED

This section describes the major independent variables in the DFGD system, both those that were evaluated and those that were held relatively constant. Also included is an overview of the test plans which were used. Several variables have been observed by others (1, 2, 3, 4) to have an influence on spray drying SO_2 removal efficiencies. These include lime stoichiometry, approach-to-saturation temperature, SDA residence time, atomizer speed, recycled solids, and inlet flue gas temperature. Test schedules prevented all of these variables from being evaluated in a single test plan. Therefore, modularized test plans were developed, each for a specific purpose. To allow sufficient and statistically adequate data to be collected in the expected short test period, test plans based on factorial and response surface designs were developed. A statistical and engineering analysis was then performed for determining the effect of the variables on the SDA, the baghouse, and the overall SO_2 removal efficiencies.

Major Variables

The lime stoichiometry is defined as moles of $Ca(OH)_2$ per mole inlet SO_2 and refers only to the fresh or makeup $Ca(OH)_2$ pumped to the SDA. This definition also applies to the more descriptive term of the external lime stoichiometry. The external lime stoichiometry was evaluated over a range from 0.7 to 1.6 mole $Ca(OH)_2$ per mole inlet SO_2 during these tests. The internal lime stoichiometry is the combination of the external (fresh) lime stoichiometry and unreacted lime stoichiometry returned to the system via the recycle slurry; it also has the units of moles $Ca(OH)_2$ per mole inlet SO_2. The internal lime stoichiometry could not be independently set during the test program since it fluctuated depending on the other process conditions. Thus the external lime stoichiometry is the more important of the two stoichiometries for process control purposes, and the term lime stoichiometry in this paper will refer specifically to the external lime stoichiometry.

The approach-to-saturation temperature is defined as the temperature difference (in F^o) between the actual flue gas temperature (dry bulb) and the flue gas wet bulb temperature as measured at the SDA exit. During the tests at the STF, the wet bulb temperature at the SDA exit varied only from 125 to 130^oF. A range of approach-to-saturation temperatures from 15 to 45 F^o was tested.

The SDA residence time is the time (in seconds) that the flue gas remains in the vessel for reaction with the atomized slurry. This time is calculated based on the SDA outlet gas conditions. Although the flue gas temperature varies throughout the vessel, the flue gas is dramatically and immediately cooled at the top of the SDA and rapidly approaches the exit gas temperature. Thus, basing the residence time on the exit gas conditions introduces little or no error into the calculation of the actual gas residence time. The design of the vessel was based on a 10-second residence time and variations from this value were investigated. To minimize the effects of poor gas distribution in the SDA (due to operation below design limits) and to remain within the limits of the ID fan, a residence time range of 8 to 12 seconds was tested.

The atomizer speed has three possible settings, depending on which motor pulley wheel was used: 12000, 14000, and 16000 rpm. All three levels were evaluated during the testing. These speeds correspond to an atomizer wheel tip speed of 390, 455, and 520 ft/sec, respectively. Although the size of the droplets sheared from the atomizer wheel was the variable of interest, the droplet size could not be measured directly. However, previous work by others (1,2,4) has determined that the droplet size is closely related to the atomizer tip speed, and thus the atomizer speed was measured as the variable of interest.

The inlet flue gas temperature was evaluated in a separate test plan. This temperature (^{o}F) is measured immediately prior to the flue gas entering the SDA with 300^{o}F being the base case condition. Temperatures of 280 and 320^{o}F were chosen as a typical range over which the temperature in a full-scale system might fluctuate.

The recycled solids rate was also evaluated in a separate test plan. Waste solids from the SDA and baghouse were slurried and recycled back to the atomizer. The recycled solids rate is reported as pounds of recycle solids per pound of fresh $Ca(OH)_2$ feed solids. Because of the water constraints imposed by the high-sulfur coal conditions, the recycled solids rate could not be varied as an independent variable over a wide range. Thus for all of the tests except the specific recycle evaluation tests, the recycled solids rate was set at its maximum value, i.e., trim water to control the approach temperature was replaced entirely with recycle slurry. In the recycle evaluation tests, fractions of the maximum recycled solids rate were used with trim water added as needed to maintain the desired approach-to-saturation temperature. These latter tests permit a separate evaluation of recycle effectiveness and could allow the effect of recycle to be incorporated into the other evaluations where necessary.

Other Important Process Parameters

The inlet SO_2 concentration was not a variable in the test plan. Instead, the SO_2 concentration in the flue gas was simply accepted as received from the boiler and typically ranged from 2500 to 3000 ppm SO_2.

Because of the water constraints in a high-sulfur coal application, the lime slurry weight percent solids was maintained at the highest possible value to minimize the amount of water needed to deliver the desired $Ca(OH)_2$ feed. (The lime slurry was pumped to the atomizer separately from the recycle slurry.) The viscosity of a lime slurry, however, increases with increasing weight percent solids and rapidly reaches a level at which it cannot be handled by conventional mixing and pumping equipment. At the STF, this upper limit on the lime slurry weight percent solids was 25 percent. Thus all of the tests were run with a lime slurry containing 24-25 percent solids.

The quality of the pebble lime can have a significant impact on the spray dryer FGD system, particularly if the grits are not removed in the slaker. For the STF tests, a highly reactive (defined by ASTM C110.2.2), high-calcium (greater than 93 percent available CaO), pebble lime was specified.

Because of the water-limited nature of this high-sulfur coal application, the recycle slurry weight percent solids are maintained as high as possible to maximize the amount of recycle per unit of water in the system. At STF, a recycle slurry percent solids of 35 percent was used to minimize operational and maintenance problems. Unlike for lime slurry, an increase in recycle solids concentration does not result in a slurry with high viscosities, but it does contribute to abrasion and plugging problems.

RESULTS

The results discussed in this section reflect the statistical analysis of the six major process design variables: (1) fresh lime stoichiometry, (2) approach-to-saturation temperature in the SDA, (3) atomizer speed, (4) SDA residence time, (5) recycled solids, and (6) inlet flue gas temperature. These analysis results indicate which major process variables are important in the SDA, the baghouse, and the overall system.

The purpose of this section is to discuss the results of the most appropriate predictive linear models of SO_2 removal efficiency that can be used for

DFGD design purposes when the above control variables are considered. Other conclusions of the analysis are also included.

Major Control Variables

The first series of tests evaluated five of the major process design variables: (1) fresh lime stoichiometry, (2) approach-to-saturation temperature in the SDA, (3) atomizer speed, (4) SDA residence time, and (5) recycled solids. The other major variable, inlet flue gas temperature, was not included in the analysis since it was held constant at 300°F for these tests. The data consist of observations from 39 tests in which all the variables were varied in a factorial test design. The recycled solids were set at the maximum possible value for each test since the recycle ratio could not be varied independently of the other four variables. Because of this experimental design limitation, lime stoichiometry of the recycle material is the only recycle parameter that is sufficiently estimable in a linear statistical model of SO_2 removal efficiency which includes the other four variables. Apparently, recycle stoichiometry is the most estimable recycle parameter because it is the least correlated with recycle ratio, whose maximum is strongly determined by the choices of lime slurry stoichiometry and approach-to-saturation temperature.

The analysis of these 39 tests indicates that the overall SO_2 removal efficiency can be accurately modeled using only two variables, the fresh lime stoichiometry and the approach-to-saturation temperature in the SDA (coefficient of determination, $R^2 = 0.95$; Mean Square Error, MSE = 9.4). The SDA SO_2 removal efficiency can be adequately modeled using the same two variables ($R^2 = 0.93$, MSE = 10.8). The baghouse SO_2 removal efficiency, however, is best represented ($R^2 = 0.85$, MSE = 55.6) by a model containing only the fresh lime stoichiometry and the recycle lime stoichiometry. However, the approach-to-saturation temperature in the SDA could be substituted for the recycle lime stoichiometry with only a slight loss of predictive capability. Therefore, the latter baghouse model (R2 = 0.82, MSE = 61.9) is used in subsequent examples since recycle lime stoichiometry is not a design test condition. It should be noted that the sum of the fresh lime and the recycle lime stoichiometries are by definition the internal lime stoichiometry. Thus, the baghouse SO_2 removal efficiency seems to be best represented as a function of only the internal lime stoichiometry. The atomizer speed and the SDA residence time variables were found to have an insignificant effect on the SO_2 removal efficiencies over the joint ranges tested once lime stoichiometry and approach-to-saturation temperature are established, assuming maximum possible recycled solids rate is used. The best model of overall SO_2 removal efficiency indicates that the SO_2 removal efficiency increases as the lime stoichiometry increases. It also increases as the approach-to-saturation temperature in the SDA decreases. Each of these variables is discussed below.

Effect of Lime Stoichiometry

The effect of fresh lime stoichiometry on the overall, the SDA, and the baghouse SO_2 removal efficiencies (as a percentage of the inlet SO_2 to the SDA) is shown in Figure 2. As can be seen, the overall SO_2 removal efficiency is highly influenced by the fresh lime stoichiometry. The same trend (i.e., increasing SO_2 removal at higher lime stoichiometries) occurs at other approach-to-saturation temperatures as shown in Figure 3a. This trend is also shown in Figures 3b and 3c where SDA SO_2 removal efficiency and baghouse SO_2 removal efficiency are shown as a function of lime stoichiometry. The baghouse efficiency in Figure 3c is expressed as percent of inlet SO_2 to the baghouse.

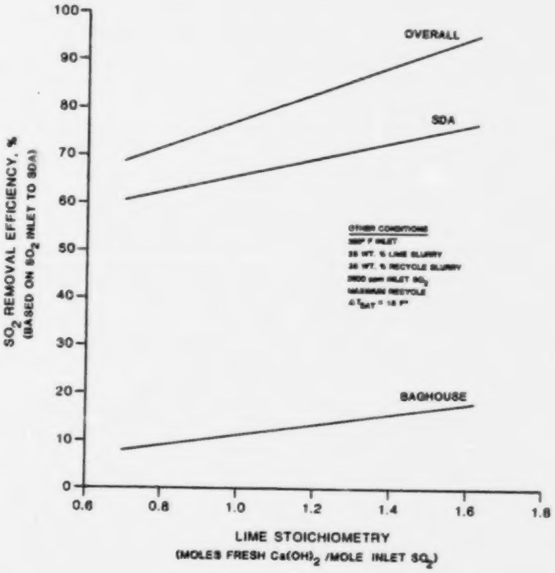
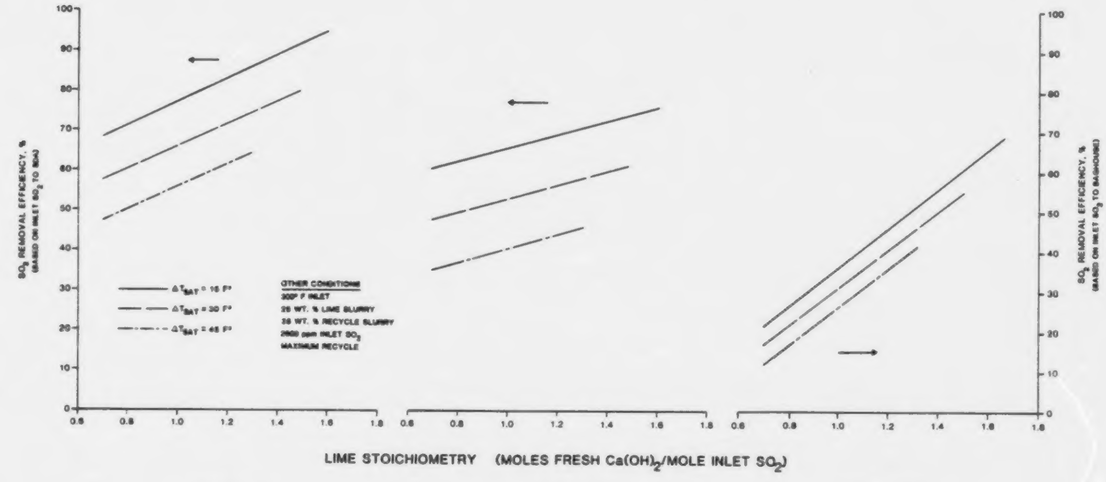


FIGURE 2. PERFORMANCE OF SYSTEM AND COMPONENTS

The most important conclusion from this data is shown in Figure 2. The SDA alone cannot achieve high SO_2 removal efficiencies (i.e., greater than 76 percent) even at high lime stoichiometries at the most optimistic (15 F°) approach-to-saturation temperature. Thus, the SO_2 removal provided by the baghouse (or other particulate control device) is essential for high-sulfur coal applications and the performance of the baghouse is crucial to the overall operation of the DFGD system.

Figure 2, however, is somewhat misleading in that the baghouse SO_2 removal efficiency is shown as a percent of the inlet SO_2 to the SDA. If the baghouse SO_2 removal efficiency is calculated based on the inlet SO_2 to the baghouse as shown in Figure 3c, it can be seen that the baghouse alone is an efficient SO_2 removal device at the higher lime stoichiometries. At an approach-to-saturation temperature of 15 F° and at lime stoichiometries greater than 1.0 mole fresh $Ca(OH)_2$/mole inlet SO_2, SO_2 removal efficiencies of 35 percent to nearly 70 percent of the inlet SO_2 to the baghouse can be achieved. The second most important conclusion from these data is that the fresh lime stoichiometry is a much more important determinant of the baghouse SO_2 removal efficiency than the SDA SO_2 removal efficiency. As shown in Figures 3b and 3c, the slope of the line in Figure 3c for the baghouse is much steeper than in Figure 3b for the SDA. The overall SO_2 removal efficiency increases faster than the SDA SO_2 removal efficiency for the same increase in lime stoichiometry to the SDA.

The reason for this larger effect of lime stoichiometry on the SO_2 removal efficiency in the baghouse is related to the effective lime stoichiometry in each vessel. As the lime stoichiometry in the SDA increases, the lime utilization in the SDA decreases dramatically (for simplicity, the lime stoichiometry in the recycle solids is neglected). For example, if a lime stoichiometry of 0.70 moles $Ca(OH)_2$/mole inlet SO_2 case is assumed, the SDA SO_2 removal efficiency is 61 percent (see Figure 2). Thus, 0.61 moles of $Ca(OH)_2$ and 0.61 moles of SO_2 react in the SDA and the effective lime stoichiometry in the baghouse is 0.23 moles of $Ca(OH)_2$/mole inlet SO_2 (0.09/0.39). A lime stoichiometry of 1.60 moles $Ca(OH)_2$/mole inlet SO_2 achieves 76 percent SO_2 removal in the SDA and the resulting effective lime



(A) OVERALL

(B) SDA

(C) BAGHOUSE

FIGURE 3. EFFECT OF STOICHIOMETRY AND APPROACH - TO - SATURATION TEMPERATURE

stoichiometry in the baghouse is 3.5 moles $Ca(OH)_2$/mole inlet SO_2 (0.84/0.24). Therefore, it is not surprising that increasing the lime stoichiometry has a more dramatic effect on the baghouse SO_2 removal efficiency.

Effect of Approach-to-Saturation Temperature

The effect of the approach-to-saturation temperature on the overall, the SDA, and the baghouse SO_2 removal efficiencies is shown in Figures 3a, 3b, and 3c, respectively. A comparison of these three figures indicates that the approach-to-saturation temperature in the SDA has a much smaller impact on the baghouse than on the SDA or overall SO_2 removal efficiency. Whereas, increasing the approach-to-saturation temperature in the SDA by 15 F° results in a 5 percentage point decrease in baghouse SO_2 removal efficiency, it results in a 12 percentage point decrease in the SDA SO_2 removal efficiency (and an 11 percentage point decrease in the overall SO_2 removal efficiency). Nevertheless, because a significant proportion of the overall SO_2 removal efficiency occurs in the SDA where the approach-to-saturation temperature is a major factor, the overall SO_2 removal efficiency is a strong function of the approach-to-saturation temperature in the SDA.

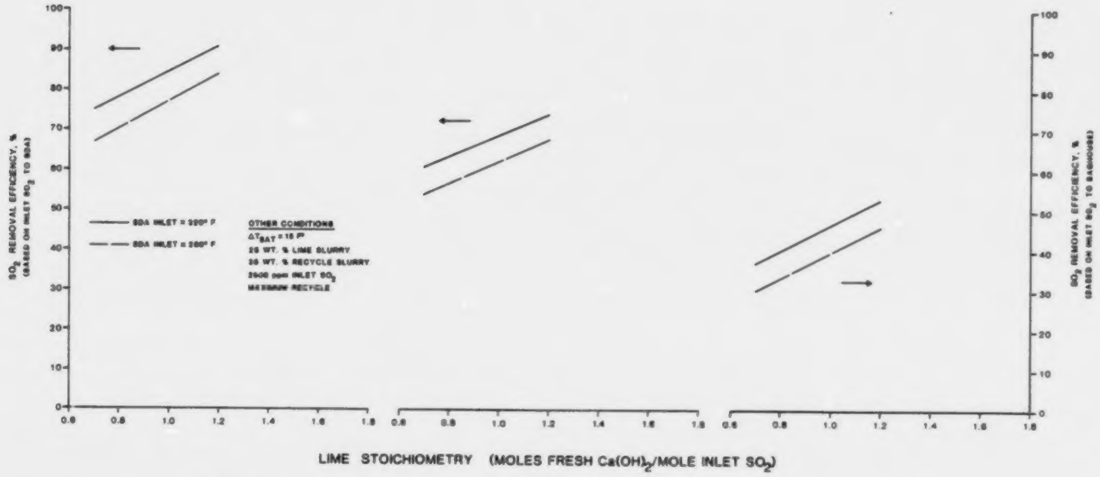
These results are not unexpected since previous work by others (1,2) has indicated that the SDA SO_2 removal efficiency is a strong function of the approach-to-saturation temperature. The reason for this strong effect is based on the following hypothesis. Since the rate controlling step in the overall reaction between SO_2 and $Ca(OH)_2$ is the absorption of SO_2 into the liquid phase, retaining liquid water in the SDA should enhance SO_2 removal. Decreasing the approach-to-saturation temperature results in a lower driving force for evaporation of water, and thus the water injected into the SDA remains in the liquid phase longer. The fact that the baghouse SO_2 removal efficiency is not affected as strongly by the approach-to-saturation temperature in the SDA is not surprising since the SDA is designed to minimize the presence of liquid water in the baghouse to avoid wet solids and bag blinding and to minimize corrosion.

Effect of Inlet Flue Gas Temperature With Other Major Variables

The next series of tests evaluated three of the major process design variables jointly: (1) fresh lime stoichiometry, (2) approach-to-saturation temperature in the SDA, and (3) the SDA inlet flue gas temperature. Recycled solids, again, were maximized for every test, and the other major variables were held constant. In this analysis, the effect of varying the inlet flue gas temperature between 280°F and 320°F was examined along with lime slurry stoichiometry and approach-to-saturation temperature. A separate standard factorial design with 11 tests was used to acquire the necessary test data.

The analysis of the data indicates that the most appropriate models for predicting the overall, the SDA, and the baghouse SO_2 removal efficiencies require the inclusion of all three variables. As was the case for the previous series of tests, the confidence in the model for the overall (R^2 = 0.95, MSE = 10.3) and SDA (R^2 = 0.98, MSE = 4.45) SO_2 removal efficiencies is much higher than that for the baghouse model (R^2 = 0.72, MSE = 46.9). These models indicate that as the lime stoichiometry and the SDA inlet temperature increase and the approach-to-saturation temperature (in the SDA) decreases within the joint ranges tested, the SO_2 removal efficiency increases.

The effect of the SDA inlet flue gas temperature on the overall, the SDA, and the baghouse SO_2 removal efficiencies is shown in Figures 4a, 4b, and 4c, respectively. The SDA inlet flue gas temperature appears to increase the SDA and the baghouse SO_2 removal efficiencies by approximately the same marginal amount once the effect of the other two variables has been accounted for. For example, increasing the SDA inlet flue gas temperature by 40°F



(A) OVERALL (B) SDA (C) BAGHOUSE

FIGURE 4. EFFECT OF STOICHIOMETRY AND SDA INLET TEMPERATURE

seems to increase the SDA and baghouse SO_2 removal efficiency about 7 percentage points, resulting in an overall SO_2 removal efficiency increase of about 8 percentage points.

Although the inlet flue gas temperature could theoretically have a direct impact on the overall SO_2 removal efficiencies (for example, by increasing the rate of reaction between lime and SO_2, etc.) the simpler and far more likely explanation is that the higher inlet flue gas temperature increases the SO_2 removal efficiency indirectly by increasing the "spray down" temperature, and thus allowing more water and hence more recycle slurry to be added to the SDA. (The impact of recycled solids on the SO_2 removal efficiency is discussed below.) Initial statistical analysis of the STF data has indicated that the increased performance at the higher inlet flue gas temperatures may be due mostly to the increased recycle since modeling of the data indicated that the SDA inlet temperature variable could be replaced by the recycle variable without much change in the predictive capability of the model. This result would not be unreasonable since previous work by others (3) has indicated that an increased recycle rate improves the drying characteristics of the feed slurry in the SDA, and the STF data indicate that an increased recycle rate improves the baghouse SO_2 removal efficiency by providing additional available alkalinity in the baghouse solids.

Effect of Recycled Solids

A third series of tests evaluated the effect of the recycled solids on the performance of the DFGD system. The first purpose of this test series was to determine if the recycle quantity is an important process design variable and if so, what parameter terminology should be used to describe the recycled solids.

Previous work by others (2,3) has indicated that the recycling of SDA and/or baghouse solids has a significant effect on SO_2 removal efficiency; however, most of this work was performed with low-sulfur, highly alkaline fly ash coals. Thus, an analysis was performed to determine if recycle is an important process design variable for high-sulfur, low-alkaline fly ash coals, and also which of five possible descriptions of the recycled solids best represent the contribution of recycle to the SO_2 removal efficiency.

The five recycle descriptions considered were the lime stoichiometry of the recycle slurry (moles $Ca(OH)_2$ in recycle slurry per mole inlet SO_2), percent solids in the feed slurry, recycle ratio (lbs recycle solids per lb fresh $Ca(OH)_2$ solids), total solids in the feed slurry (lbs per minute), and total solids in the recycle slurry (lbs per minute). The data used in the analysis are based on 36 tests using both maximum possible and submaximum recycled solids with a wide range of recycle slurry percent solids to enhance estimability of each recycle parameter. Other key factors except lime stoichiometry were controlled at baseline conditions. The analysis first accounted for the presence of lime stoichiometry as a contributor to the SO_2 removal efficiency in the overall system, the SDA, and the baghouse. Then the additional effects of each recycle parameter as a second explanatory variable were compared.

The analysis of the data indicates that even for a low alkaline ash the recycled solids are an important determinant of the overall, the SDA, and the baghouse SO_2 removal efficiency. Thus, some recycle parameter should be included in any model of the SO_2 removal efficiency in a DFGD system where maximum possible recycled solids are not always used. Although all five recycle variables were significant contributors to SO_2 removal efficiency, some were stronger determinants than others. In the SDA, the feed slurry percent solids were most appropriate, but the total feed solids and the total recycle feed solids were only slightly less significant. It has been hypothesized that inreasing the solids content of the feed slurry increases the slurry volume which results in more slurry droplets, thus a higher droplet surface area. This increased droplet surface area promotes the initial SO_2 absorption reaction with water with a subsequent increase in SO_2 removal. In the baghouse, the recycle lime stoichiometry showed the strongest relationship with the SO_2 removal efficiency while feed slurry percent solids, although second most significant, was relatively poor in comparison with recycle lime stoichiometry. Since the droplet surface area effect is no longer applicable in the baghouse, the additional alkalinity from the recycled solids becomes the dominant factor promoting additional SO_2 removal with recycle. For the overall SO_2 removal efficiency, the feed slurry percent solids were the most appropriate parameter (with total feed solids and total recycle solids a relatively poor second and third, respectively). Since the performance in the SDA is the dominant factor in the performance of the system, the feed slurry percent solids describing the contribution of recycle on the system is expected.

FULL-SCALE CONSIDERATIONS

Although the pilot-plant data have indicated the relative importance of the major process design variables, in the design of a full-scale DFGD system other considerations (i.e., process economics, operating and maintenance problems, etc.) may dictate the variable levels. Each of the major process variables is discussed in more detail below.

Process Design Variables

Based on the results from these initial tests with the R-C pilot plant, it is apparent that several of the six primary variables evaluated in these initial tests have a limited effect on the system performance (i.e., SO_2 removal efficiency). For example, the data indicate that the most important consideration when evaluating the SDA residence time and the atomizer speed is whether these variables are in the right range to prevent the production of wet solids, since minor (± 10 percent) changes in either variable can be made with no apparent effect on system performance. Thus, in a full-scale design, the SDA design residence time is set at a level that ensures an adequate residence time at maximum load. The atomizer is then designed to provide the optimum speed at this residence time.

The remaining major process variables (lime stoichiometry, approach-to-saturation temperature, inlet flue gas temperature, and recycled solids) are somewhat interrelated and a more detailed discussion of this interrelationship is needed before discussing the full-scale consideration of each of them. The primary technical constraint in a DFGD system is the amount of water which can be evaporated in the SDA. Since no external heat source is provided, only the heat content of the flue gas is available to evaporate the water. Thus, the most important variables in the DFGD system are the inlet flue gas temperature (which determines the heat content of the flue gas to the DFGD system) and the approach-to-saturation temperature (which determines the heat content of the flue gas at the SDA exit). These variables determine the total flow of water to the SDA. Typically, the inlet flue gas temperature is determined when the boiler is specified, and thus cannot be arbitrarily changed to accommodate the DFGD system. The approach-to-saturation temperature is also usually specified in order to maximize the SO_2 removal efficiency in the DFGD system while maintaining system operability and minimizing corrosion in the downstream equipment. Thus, in a full-scale DFGD system, the inlet flue gas temperature and the approach-to-saturation temperature, and hence the amount of water which can be evaporated in the SDA (at full load), are fixed and cannot be manipulated.

All of the major DFGD variables have an impact on this water balance constraint in one way or another. In a high-sulfur coal application, these relationships are much more important than in a low-sulfur coal application where, because of the lower sulfur levels, the system designers have much more flexibility in designing the system. Each of these variables is discussed in more detail below.

Approach-to-Saturation Temperature--The approach-to-saturation temperature in the SDA is one of the two most important variables affecting the overall system performance (the fresh lime stoichiometry is the second). The approach-to-saturation temperature in the SDA is considered the more important of the two variables because it is the only process design variable which can be easily changed and can have a dramatic effect on the water balance and also because it has a significant effect on the overall SO_2 removal efficiency in the DFGD system.

Although the heat available in the flue gas is determined by the flue gas temperature, the amount of this heat which can be used to evaporate water is determined by the approach-to-saturation temperature at the SDA exit. The flue gas adiabatic saturation temperature is determined by the flue gas conditions (i.e., the temperature and pressure) and to a lesser extent, its composition. The approach-to-saturation temperature indirectly specifies the SDA outlet temperature; thus, the amount of water which can be evaporated in the SDA is determined. Operating the DFGD system at a lower approach-to-saturation temperature means more water can be injected into the SDA. Conversely, operating the DFGD system at a higher approach-to-saturation temperature means less water can be injected into the SDA in the form of alkaline slurry.

The second consideration when determining the approach-to-saturation temperature is its impact on the overall SO_2 removal efficiency in the DFGD system. Since the presence of liquid water is necessary for the SO_2 removal reaction, the longer liquid water is present in the SDA, the higher the SO_2 removal efficiency. A lower or closer approach-to-saturation temperature both increases the amount of water injected into the SDA and decreases the driving force for the evaporation of the last water in the solids (i.e., the temperature difference). This increases the time that liquid water is present and, thereby, increases the SO_2 removal efficiency in the SDA. Thus, the SDA should be operated as close to the saturation temperature as practical without producing wet solids or endangering downstream equipment from moisture condensation and subsequent corrosion. Based on the TVA tests, this closest approach-to-saturation temperature for high-sulfur coal applications is 18-20 F°.

Lime Stoichiometry--The fresh (or external) lime stoichiometry is the second of the two most important variables affecting the overall system performance. Because of the unusually rapid increase in viscosity of a lime slurry as the weight percent solids in the lime slurry increases, these slurries cannot be mixed or pumped with conventional equipment. This technical limitation on the lime slurry being fed to the SDA is typically reached at 24-25 percent solids.

In low-sulfur coal applications, lime stoichiometries substantially below 1.0 mole $Ca(OH)_2$/mole inlet SO_2 have been reported; however, in high-sulfur coal applications, lime stoichiometries greater than 1.0 mole $Ca(OH)_2$/mole inlet SO_2 are typical. This 30-40 percent increase in the lime stoichiometry means more water per mole of inlet SO_2 is used to transport the lime to the SDA. A more important consideration in water requirements is that the stoichiometry in the DFGD system is based on total moles of inlet SO_2, and thus at a constant lime stoichiometry of 1.0 mole $Ca(OH)_2$/mole inlet SO_2, the lime slurry flow rate (and hence water consumption to transport lime) for a high-sulfur coal application will be substantially higher than a low-sulfur coal application because of the increase in the total moles of inlet SO_2. For a 3.5 percent sulfur coal case, the lime slurry feed rate will be about 3.5 times that for a 1.0 percent sulfur coal case at the same lime stoichiometry (for simplicity, the increase in lime stoichiometry due to the high inlet SO_2 levels has been neglected).

Thus, in contrast to a low-sulfur coal application, a large fraction of the water entering the SDA in the high-sulfur coal application is used to transport the lime. The fraction of the total water used to transport lime will vary significantly based on boiler conditions, coal sulfur level, and the required lime stoichiometry. However, since the lime stoichiometry is one of the most important variables, the amount of water required to transport this lime to the spray dryer is allocated first from the total amount of water to be evaporated in the SDA. After the amount of water to transport the lime is subtracted, any remaining water is used for reslurrying the FGD waste to be recycled. For a high-sulfur coal application, about 50-80 percent of the total water to the SDA is used to transport lime at full boiler load. At lower boiler loads when the inlet flue gas temperature drops, the percentage of water used to transport the lime increases dramatically.

Inlet Flue Gas Temperature--The inlet flue gas temperature to the SDA is important since it determines the heat which is available to evaporate water. The higher the inlet temperature, the more the heat, and hence the more water which can be evaporated in the SDA. Normally in a full-scale system, however, the inlet flue gas temperature is determined by the design and operation of the boiler and cannot be controlled. The inlet flue gas temperature at full load will, however, vary with significant changes in the ambient temperature as in seasonal variations.

In a full-scale application, the inlet flue gas temperature is an important consideration since during boiler turndown the inlet temperature tends to decrease, and thus have a direct impact on the DFGD system. As with the full-load situation, the inlet flue gas temperature at reduced loads is determined by the boiler design and operating conditions, and thus the DFGD system must be designed to accommodate boiler turndown conditions. The effect of decreasing the inlet flue gas temperature at reduced boiler loads is to reduce the amount of water which can be evaporated in the SDA. Since the lime stoichiometry and the approach-to-saturation temperature cannot be changed, the only way to reduce the water flow rate to the SDA and compensate for the lower inlet flue gas temperature is to reduce the recycle slurry flow rate. Boiler operating scenarios must be evaluated in terms of its potential effect on a proposed DFGD system.

Recycle Ratio--The recycle ratio is a measure of the amount of FGD waste which can be recycled to increase lime utilization. Since the recycle ratio is of lesser importance than the lime stoichiometry or the approach-to-saturation temperature, the amount of FGD waste which can be recycled is determined by water balance considerations, and thus is not an independent variable (i.e., it cannot be specified). The other major process variables, which determine the amount of water which can be added, are optimized first and any remaining water is used to reslurry the waste material for recycle.

For low-sulfur coal applications where the proportion of the total available water required to transport the lime in slurry form to the SDA is relatively small, large recycle ratios (8 to 12 lbs of recycle solids per lb of fresh $Ca(OH)_2$ feed solids) are possible. For high-sulfur coal applications where a substantial portion of the total available water is required to transport the lime to the SDA, recycle ratios from 0.0 to 3.0 are more typical. At reduced loads where the inlet flue gas temperature drops, the recycle ratio which can be achieved may decline to values near zero.

The data from the STF seems to indicate that maximizing the recycle ratio by increasing the weight percent solids in the recycle slurry enhances the SO_2 removal efficiency in the DFGD system. However, the weight percent solids in the recycle slurry is limited to the range of 35 to 40 percent because of the potential for erosion in the recycle system.

Implications for Fuel Selection

The selection of a DFGD system for a high-sulfur coal application requires a detailed determination of the overall SO_2 removal efficiency required for the unit based on the coal sulfur level, the emission regulation, and type of boiler operation (e.g., base load, cycling, etc.). One cannot simply assume that the design of the DFGD system can be modified to achieve the required SO_2 removal efficiency since the DFGD system is not as flexible as a wet FGD system for high-sulfur coal applications.

Because of the technical constraints imposed on the SDA by the water balance and the pumpability of the lime slurry, the DFGD system must be evaluated under the worst-case conditions to determine if the DFGD system is technically and economically feasible for a particular high-sulfur coal application. These worst-case conditions will typically occur at reduced boiler load and low ambient temperatures, the combination of which determines the lowest inlet flue gas temperature. This lowest inlet flue gas temperature in combination with the lowest flue gas rate (also at reduced boiler load) defines the lowest addition rate for water into the SDA. (The approach-to-saturation temperature remains constant at 20 F°.) The minimum water addition rate defines the maximum amount of lime (and hence the lime stoichiometry) which can be added to the SDA because of the limitations on the weight percent solids in the lime slurry. This maximum lime stoichiometry (in combination with the already defined inlet flue gas temperature and approach-to-saturation temperature) defines the maximum overall SO_2 removal efficiency in the system. Thus, knowing the maximum SO_2 removal efficiency achievable in the system and the SO_2 emission regulation defines the maximum allowable sulfur level in the coal.

SUMMARY AND CONCLUSIONS

TVA, in cooperation with R-C, has investigated the application of the spray drying FGD technology to the treatment of flue gas from burning high-sulfur coal. The results of this high-sulfur testing show trends similar to those observed by others when testing on low-sulfur, coal-derived flue gas (i.e., either increasing the lime stoichiometry or decreasing the approach-to-saturation temperature increases the SO_2 removal efficiency). These variables are the key parameters in designing a DFGD system. Relative to these two parameters, flue gas residence time and atomizer speed were found to have a negligible effect on SO_2 removal efficiency over the range tested. However, both are important in ensuring that a dry FGD waste is produced. The inlet flue gas temperature has a significant effect on the system performance (increasing the inlet flue gas temperature increases the SO_2 removal efficiency). Higher recycling of solids increases both lime utilization and the SO_2 removal efficiency. However, the recycle rate is limited by the water balance in the SDA and is impacted by the control variables (i.e., lime stoichiometry, approach-to-saturation temperature, and inlet flue gas temperature), and the physical limitations of the equipment (i.e., pumpability and erosion of the recycle slurry).

In addition to the above, the water balance is of major concern in a full-scale high-sulfur coal DFGD application. In contrast to low-sulfur coal applications, the water balance in high-sulfur coal applications becomes the limiting technical constraint. Since the inlet flue gas temperature is determined by the boiler design and the approach-to-saturation temperature is determined by the necessity to remove as much SO_2 as possible, the maximum lime stoichiometry and hence SO_2 removal are defined by the SDA water balance. The lime stoichiometry is determined by the SO_2 removal efficiency requirements and any remaining water is used to recycle the collected FGD waste. Typical boiler operating ranges also affect the recycle capability. Higher than expected inlet SO_2 levels and/or lower inlet temperature result in lower than expected SO_2 removal efficiencies. Thus, the DFGD system for a high-sulfur coal application must be designed for the worst case water balance conditions. From the data generated through testing at STF, TVA has determined that the DFGD process is technically feasible for a high-sulfur (4.5% dry basis) coal application where the emission regulation is 1.2 lb SO_2/MBtu (an 83 percent SO_2 removal efficiency) and under SDA inlet conditions similar to those evaluated at STF. This application, however, is approaching the upper limit for this technology and is subject to a site-specific evaluation.

ACKNOWLEDGEMENT

The authors wish to acknowledge the efforts of Steve Newton and the operating and maintenance personnel at the Shawnee Test Facility in performing the testing for this evaluation.

NOTICE

This publication was prepared by the Tennessee Valley Authority (TVA) as an account of work sponsored by TVA and Research-Cottrell (R-C). Neither TVA, nor persons acting in behalf of TVA or R-C: (a) makes any warranty or representation, express or implied, with respect to the accuracy, completeness, or usefulness of the information contained in this report, or that the use of any information, apparatus, method, or process disclosed in this report may not infringe privately owned rights; or (b) assumes any liabilities with respect to the use of, or for damages resulting from the use of, any information, apparatus, method, or process disclosed in this publication. This paper does not necessarily reflect the views and policies of TVA.

REFERENCES

1. Apple, C. and M. E. Kelly, Mechanisms of Dry SO_2 Control Processes, EPA-600/7-82-026, April 1982.

2. Parsons, E. L., Jr., et al., "SO_2 Removal by Dry FGD," Proceedings of Symposium on Flue Gas Desulfurization, EPA-600/9-81-091b, April 1981.

3. Blythe, G. M. and R. G. Rhudy, "EPRI Spray Dryer/Baghouse Pilot Plant Status and Results," Proceedings: Eighth Symposium on Flue Gas Desulfurization, EPRI CS-3706, Volume 2, November 1984.

4. Samuel, E. A., et al., "Process Characterization of SO_2 Removal in Spray Absorber/Baghouse Systems," Proceedings: Eighth Symposium on Flue Gas Desulfurizaton, EPRI CS-3706, Volume 2, November 1984.

RECENT RESULTS FROM THE EPRI 2-1/2 MW SPRAY DRYER PILOT PLANT

Richard G. Rhudy
Electric Power Research Institute
Palo Alto, CA 94303

Gary M. Blythe
Radian Corporation
Austin, TX 78766

ABSTRACT

Since January 1984, testing on the 2-1/2 MW spray dryer/fabric filter FGD pilot plant has centered on variations to the basic FGD system. Results which are presented in this paper demonstrate the effects of the use of several cooling tower blowdown waters for lime slaking and/or process dilution, the effects of calcium chloride as a system performance additive, and the impacts on SO_2 removal of the use of warm (spray dryer inlet) gas bypass around the spray dryer to reheat the fabric filter inlet gas by 20°F to 40°F.

Additionally, results are presented which relate process variations to fabric filter performance, such as corrosion tendencies and tubesheet pressure drop. Data are presented which compare the tube sheet pressure drop history for bags in the fabric filter compartment downstream of the spray dryer to that for bags in a control compartment that is filtering hot, untreated flue gas. Finally, data are presented which relate these process variations to the disposal properties of solids collected in the fabric filter.

INTRODUCTION

The Electric Power Research Institute has been conducting a pilot-scale study of spray dryer/fabric filter FGD technology at their Arapahoe test facility since 1982. Previous papers[1,2,3] have reported results from these studies. Most of the tests for which results were presented in these earlier papers were aimed at establishing the basic operating characteristics of the FGD system.

This paper focuses on results of more recent tests, which have evaluated variations to the basic process. Variations discussed in this paper include the use of power plant wastewater streams as makeup to the FGD system, the use of calcium chloride as an SO_2 removal performance additive, and warm gas (air heater exit) bypass around the spray dryer to reheat fabric filter inlet gas and to provide corresponding protection from corrosion. Data are presented which demonstrate the impacts of these process variations on system SO_2 removal performance, fabric filter tubesheet pressure drop performance, and system solid waste disposal properties.

Also included in this paper is an update on fabric filter corrosion in the pilot system. The update discusses the effectiveness of several wall coatings, and of improved improved fabric filter insulation in preventing moisture condensation on compartment walls.

BACKGROUND

Figure 1 is a process flow diagram for the 2-1/2 MW spray dryer/fabric filter pilot unit. The pilot unit is normally configured for rotary atomization and side gas exit from from the dryer vessel. The single fabric filter compartment is cleaned by reverse gas, normally once every three hours. The pilot system treats flue gas from a low sulfur coal fired boiler. The inlet SO_2 level is varied from 400 ppm to 3000 ppm by spiking SO_2 into the spray dryer inlet flue gas. The system inlet temperature can be controlled between approximately 270°F and 325°F. Most testing has been conducted with a high reactivity quicklime slaked in a paste slaker, and with recycle of fabric filter solids.

The pilot unit has two large tanks available for mixing alternate makeup waters or for storing waters trucked in from other sites. Wastewaters from these tanks can be used in the lime slurry preparation area for slaking and/or for slaker product dilution, and in the atomizer feed preparation area for feed slurry dilution and/or temperature control water. These tanks and piping systems were used to introduce simulated and actual power plant wastewater streams as makeup for the FGD system, and to introduce calcium chloride solutions used as a performance additive.

The pilot system also has provisions for warm-gas-bypass reheat around the spray dryer vessel. The gas bypassed is at the same temperature and SO_2 concentration as the spray dryer inlet gas. In the bypass reheat tests discussed in this paper, the flow of this stream was automatically controlled

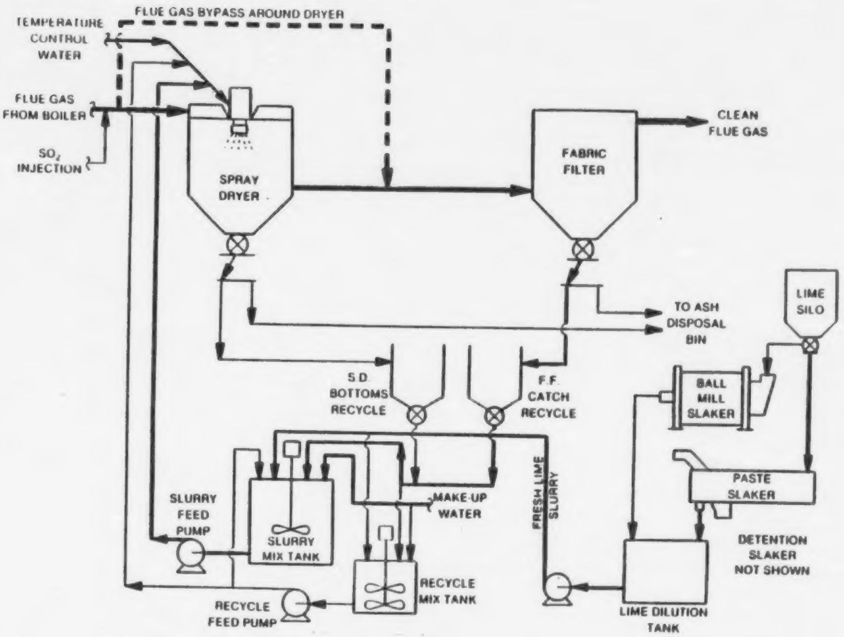

FLUE GAS BYPASS AROUND DRYER
TEMPERATURE CONTROL WATER
FLUE GAS FROM BOILER
SO_2 INJECTION
SPRAY DRYER
FABRIC FILTER
CLEAN FLUE GAS
TO ASH DISPOSAL BIN
LIME SILO
S.D. BOTTOMS RECYCLE
F.F. CATCH RECYCLE
BALL MILL SLAKER
PASTE SLAKER
DETENTION SLAKER NOT SHOWN
FRESH LIME SLURRY
MAKE-UP WATER
SLURRY FEED PUMP
SLURRY MIX TANK
RECYCLE MIX TANK
RECYCLE FEED PUMP
LIME DILUTION TANK

Figure 1. Simplified process flow diagram for spray dryer pilot plant.

to provide a preset, constant number of degrees of reheat above the spray dryer outlet gas temperature.

WASTEWATER AS FGD MAKEUP

The first series of tests to be discussed involved quantifying the impacts of using power plant wastewater streams as makeup water for the spray dryer/fabric filter FGD system. These tests concentrated on cooling tower blowdown (CTB) waters, as these represent a relatively large volume waste stream that must be reused to allow closed loop operation at a power plant. These waters were used in some tests for all of the makeup to the FGD system, including lime slaking water. In the remaining tests, fresh water was used for lime slaking and CTB waters were used for the remainder of the water entering the system.

In all of these tests, the pilot unit was operated at a 1000 ppm inlet SO_2 level, a 280°F inlet temperature, a 20°F approach to adiabatic saturation at the spray dryer outlet, and a 2:1 recycle ratio. For these conditions, lime slaking water represents about 20 to 25 percent of the system makeup water. The remainder of the makeup water is used for atomizer feed slurry batch dilution and/or for temperature control water. For lower inlet SO_2 levels, lime slaking water represents an even smaller fraction of the total system makeup. Conversely, at higher inlet SO_2 levels, lime slaking water may represent half or more of the system makeup water.

Table 1 summarizes the characteristics of the CTB waters tested. The waters chosen represent a wide variety of recirculating water system types, including gypsum-limited, acid treated systems, side stream softened systems, and silica limited systems.

SO_2 Removal Impacts

CTB1 was chosen to simulate a 1000 ppm sulfate content gypsum-limited system. This sulfate level is generally accepted as being a threshold value for lime slaking, above which the sulfate content has negative impacts on the reactivity of the product lime slurries. The results of the CTB1 tests were presented in a previous paper[2]. In these tests, rather than observing any negative impacts on system SO_2 removal performance as expected, SO_2 removal was instead enhanced. This enhancement was attributed to the high chloride content of that water.

The high chloride content of CTB1 (nominally 2000 ppm) resulted from the choice of chemicals that were mixed to achieve the 1000 ppm sulfate level and result in a high gypsum relative saturation level: calcium chloride and sodium sulfate. The performance improving effects of the high chloride levels, which were not understood at the time of these tests, appeared to overwhelm any potentially negative impacts of the 1000 ppm sulfate level even when using this simulated CTB for slaking. The effects of chloride as a performance additive are documented later in this paper. It should be noted that while the chloride level in this water (2000 ppm) is high relative to the concentrations of the other species present, it is within the range of that potentially encountered in cooling water systems.

CTB2 was selected to simulate a blowdown from a side-stream softened cooling water system. As such, the predominant ionic species present are sodium (1800 ppm) and sulfate (5000 ppm).

The results of these tests are plotted in Figure 2. The results are plotted relative to 95 percent confidence intervals for baseline data. They show that when used for all system makeup except lime slaking, CTB2 had a slightly positive effect on SO_2 removal. This positive effect is presumed to be due to the presence of relatively high sodium levels in the water. The high sodium levels could have two beneficial effects. First, high sodium levels could result in increased liquid phase alkalinity in the atomizer feed slurry. This would enhance initial SO_2 removal rates as the droplets are atomized into the SO_2-laden flue gas. Second, because of the high water solubility of most sodium-based salts, these salts are more difficult to dry than corresponding calcium salts. Thus, the presence of high sodium levels in the atomizer feed slurry may cause droplets to dry more slowly. Increased droplet residence time would also increase the time available for liquid phase SO_2 removal reactions. The presence of harder to dry sodium salts may also result in higher residual moisture levels in the solids leaving the spray dryer, and thus collected in the fabric filter. Higher residual moisture levels have been observed to improve SO_2 removal efficiency across a fabric filter solids cake[4].

TABLE 1. COOLING TOWER BLOWDOWN WATERS TESTED AT ARAPAHOE

CTB Water	CTB1	CTB2	CTB3	Craig	CTB4	Cunningham
Source	Simulated	Simulated	Simulated**	Colorado Ute Craig Sta.	Simulated	Southwestern*** Public Service Cunningham Sta.
Type	Gypsum Limited	Side Stream Softened, Gypsum Limited	Acid Treated Gypsum Limited	Acid Treated Gypsum Limited	Silica Limited	Silica Limited
Analysis (mg/1)						
Ca^{++}	1160	410	420	350	182	300
Mg^{++}	-	53	220	196	27	41
Na^{+}	480	1830	490	371	230	120
$SO_4^{=}$	1000	5080	2450	2014	660	735
Cl^{-}	2050	170	157	166	120	110
$CO_3^{=}$	-	-	95	NA*	NA	NA
SiO_2	11	11	11	5	128	140
Inhibitors	No	Scale Inhib.	Scale, Corr. Inhibitor	Scale, Corr. Inhibitor	None	None
Gypsum Relative Saturation	0.75	0.90	0.80	0.65	0.21	0.38

* NA - Analysis not available.
** Simulates Craig CTB.
*** Simulates Cunningham CTB.

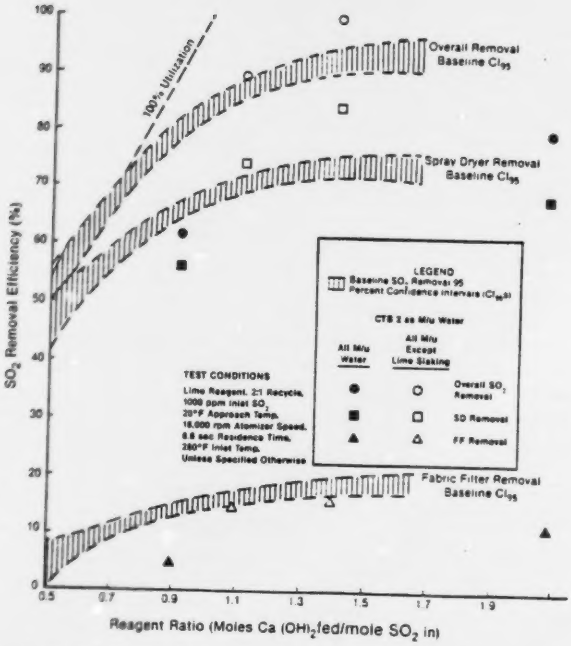


Figure 2. Results for tests with CTB2 for all system makeup water, and for all system makeup water except lime slaking, compared to baseline test 95 percent confidence intervals.

When used for lime slaking as well as for dilution, CTB2 showed very negative effects. At a reagent ratio of 0.9 moles of fresh lime per mole of SO_2 in the inlet gas, overall SO_2 removal was measured at only 61 percent, whereas baseline performance was about 80 percent removal. At an extremely high reagent ratio of 2.2 moles of lime per mole of inlet SO_2, overall SO_2 removal was limited to approximately 80 percent. Although no baseline data are available at this high reagent ratio, SO_2 removals near 100 percent would be expected.

The negative impacts of using this water for slaking have been attributed to the 5000 ppm sulfate content. The lime slurry produced during these tests was observed to be very coarse. In fact, the ball mill slaker had to be used in the second (high reagent ratio) test because the extremely coarse lime solids produced were found to collect in and plug the slaking chamber of the paste slaker. These tests confirm that high sulfate content waters can be used for lime slaking only at a penalty in product lime utilization.

CTB3 simulates a gypsum-relative-saturation-limited, acid treated cooling water system, and was chosen to duplicate the cooling water at the Colorado Ute Craig station. The results of the simulated CTB water tests were then compared to performance with the actual water. As seen in Table 1, the simulated CTB3 was slightly more concentrated with respect to sulfate content and gypsum relative saturation than the actual Craig CTB water. This discrepancy is within the normal variation in full scale cooling water system operations, and resulted because the simulated water composition was matched to a previous analysis of the Craig CTB water.

Figure 3 presents the results of the use of CTB3 and the actual Craig CTB for all makeup except lime slaking, compared to baseline 95 percent confidence intervals. Both waters show essentially no effect on overall SO_2 removal or on relative spray dryer/ fabric filter contributions to SO_2 removal.

Figure 4 shows the results of tests using CTB3 and Craig CTB for slaking as well, compared to baseline (service water only) 95 percent confidence intervals. In this case, both the simulated and actual waters showed negative impacts. This negative impact is presumably a result of the moderately high sulfate level in these waters (2000 to 2500 ppm). However, note that the actual Craig CTB water shows a more negative effect than the simulated water, in spite of having the lower sulfate content (2000 ppm vs. 2500 ppm). A measurable difference was observed when using the simulated versus actual water for lime slaking in spite of efforts to match the concentrations of major ionic species in the waters. Even corrosion and scale inhibitors were added to the simulated water to match levels in the actual water.

In an attempt to identify a trace level species that might account for the difference in performance between these two waters, trace element analyses of the species present were conducted. Table 2 compares a trace element analysis of the Craig cooling water to that of the service water used to mix the CTB3. No large differences are observed, although with no experimental data available, it is difficult to predict the impacts of variations in the concentrations of these species. It remains unclear as to why the actual water had a more negative impact than the simulated water.

CTB4 represents a silica limited cooling water system. The composition was chosen to match the cooling water composition from the Southwestern Public Service Company Cunningham station. Because this water has a relatively low sulfate content (less than 1000 ppm) it was used for slaking and for dilution in all tests. Separate dilution-only tests did not appear warranted.

The results of the tests using CTB4 and Cunningham CTB for slaking and dilution water are compared to baseline 95 percent confidence intervals in Figure 5. CTB4 shows no effect relative to baseline tests using service water for slaking and dilution. The Cunningham CTB appears to slightly improve SO_2 removal relative to baseline performance. Considering that only two tests were conducted, this difference may just reflect normal experimental scatter. As with the comparison of the Craig CTB with service water, a similar trace element comparison with Cunningham CTB in Table 3 shows no obvious differences.

It remains unclear as to why the performance of the two simulated CTB waters do not agree with that of the actual waters they simulate. In both cases, though, the use of the simulated water as system makeup results in the same directional effects (i.e., improvement versus decrease in SO_2 removal) as that of the actual water. Thus, while simulated CTB waters may not exactly predict the performance of an actual CTB water, they may be

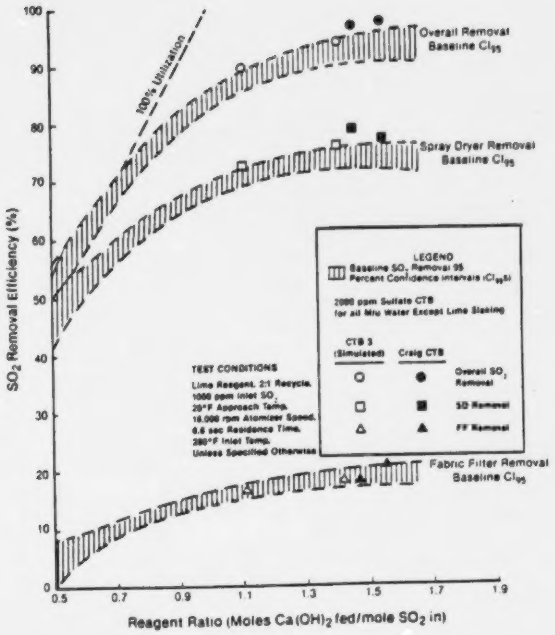


Figure 3. Results for tests with CTB3 and Craig CTB for all system makeup water except lime slaking, compared to baseline test 95 percent confidence intervals.

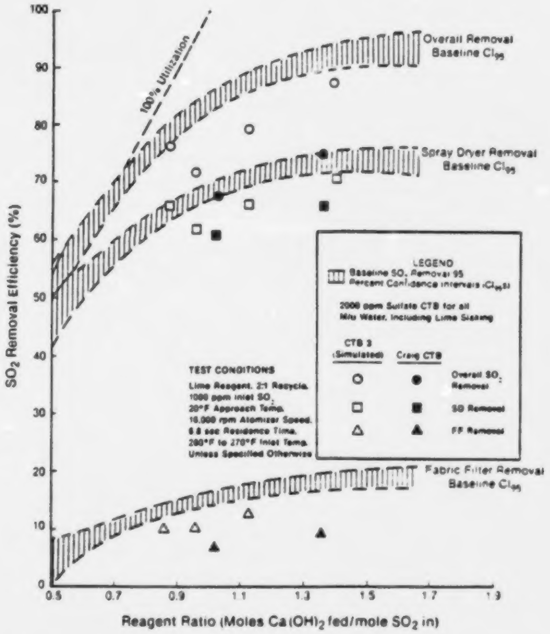


Figure 4. Results for tests with CTB3 and Craig CTB for all system makeup water, including lime slaking, compared to baseline test 95 percent confidence intervals.

TABLE 2. CTB WATER TRACE METAL ANALYSIS

Metal	Service Water, mg/l	Craig CTB, mg/l
Ag	0.005	<0.002
Al	1.1	2.1
As	<0.006	0.15
B	0.009	<0.009
Ba	0.03	0.29
Be	0.004	0.002
Cd	<0.002	<0.002
Co	<0.006	<0.006
Cr	<0.001	<0.001
Cu	0.007	0.064
Fe	0.18	0.038
K	4.2	10
Mn	0.032	<0.001
Mo	0.008	0.037
Ni	0.004	<0.003
Pb	<0.084	<0.084
Sb	<0.032	<0.032
Se	<0.084	0.14
Si	5.0	2.1
Tl	<0.09	<0.09
V	0.23	1.4
Zn	0.025	0.08

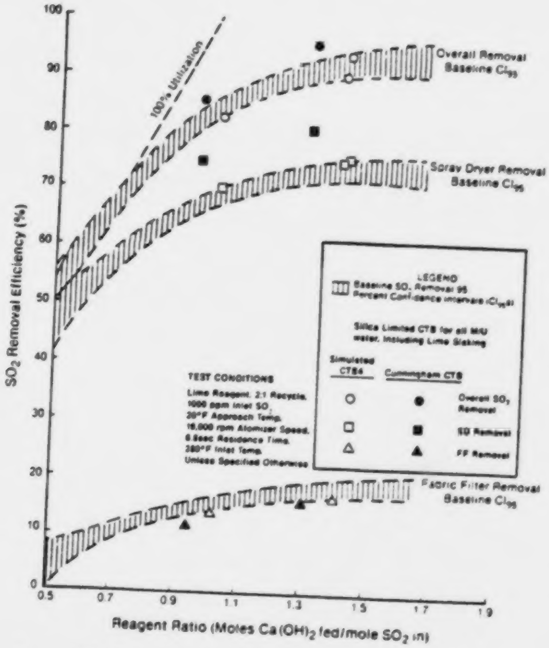


Figure 5. Results for CTB4 and Cunningham CTB for all system makeup water, including lime slaking, compared to baseline test 95 percent confidence intervals.

TABLE 3. CTB WATER TRACE METAL CONCENTRATIONS

Metal	Service Water, mg/l	Cunningham CTB, mg/l
Ag	0.005	<0.002
Al	1.1	1.3
As	<0.057	0.057
B	0.009	<0.009
Ba	0.031	0.21
Be	0.004	0.003
Cd	<0.002	<0.002
Co	<0.006	<0.006
Cr	<0.001	<0.003
Cu	0.007	0.001
Fe	0.18	0.068
K	4.2	10
Mn	0.032	<0.001
Mo	0.008	0.057
Ni	0.004	<0.080
Pb	<0.084	<0.084
Sb	<0.032	<0.032
Se	<0.084	0.092
Tl	<0.09	<0.09
V	0.23	0.38
Zn	0.025	0.022

useful in predicting whether an anticipated blowdown water source will have positive or negative impacts on system SO_2 removal performance.

Waste Property Impacts

Another potential use of power plant waste waters is for wetting of solid wastes produced by the dry FGD system. The flyash and calcium/sulfur salts are generally wetted to avoid dusting and to promote cementation reactions. The quantities of water involved vary from only a few percent where only dust control is desired to as high as 30 to 40 percent moisture added for optimum compacted strength. One question regarding solid waste handling is whether the quality of water used for wetting affects the cured waste properties: i.e., do high dissolved solids contents such as in CTB waters interfere with cementation reactions in the wetted solids? Part of this program has included a laboratory scale determination of properties of solid wastes produced by the spray dryer/baghouse system. This laboratory study has allowed direct comparison to properties of wastes wetted with deionized water with those wetted with CTB waters. Some of these results are compared here.

One step in the laboratory study of solid waste properties is to wet the solids to varying moisture levels, compact the wetted solids, then allow the samples to cure at controlled temperature and humidity for up to 56 days. These cured samples are then tested for strength and for the permeability of water through the solids. The laboratory strength tests are considered to be analogous to full-scale disposal practices, although full-scale systems are of course not subjected to exactly the same carefully controlled conditions. For example, the laboratory wetting is analogous to pug mill wetting of dry waste solids, compacting is analogous to operating heavy equipment on a landfill, etc. The strength test, then, is an indicator of how well the solids set up to allow heavy equipment to be operated on the landfill. Permeability tests provide a relative indication of how readily rainwater would soak through the cured solids and might therefore leach ionic species out of the landfill.

Table 4 summarizes unconfined compressive strength data for waste samples from a number of pilot unit test conditions. Included in the table are key test conditions, and the highest strength values measured for both deionized water and CTB waters used to wet the samples. Also listed is the percent moisture in the sample that corresponded to these highest strength levels. Table 4 shows that using CTB water rather than deionized water for wetting the solids had no significant effect on cured sample unconfined compressive strength. The first, third, and fourth samples listed showed very high values regardless of water type. As an example of how high these values appear, a person walking exerts about 3 to 5 psi, a pneumatic tire equipped truck exerts about 50 to 60 psi. All three of these samples should easily support these activities, regardless of what water quality is used for wetting.

TABLE 4. WASTE SAMPLE UNCONFINED COMPRESSIVE STRENGTH DATA

SAMPLE DESCRIPTION				
Sample No.	1	2	3	4
Inlet SO_2 Level, ppm	400	1000	1000	2000
Recycle Ratio	12:1	2.1:1	1.7:1	1.4:1
Reagent Ratio	0.9	1.1	1.6	1.3
Overall SO_2 Removal, %	92	98	98	92
Reagent Utilization, %	98	89	62	71
Dilution Water Used in FGD System	Service	CTB2	Service	Service
WASTE PROPERTIES, DEIONIZED WATER FOR WETTING				
Moisture Level for Highest Strength Value, wt. percent	18	24	21	30
Highest 56-day Unconfined Compressive Strength Value, psi	146	40	156	293
WASTE PROPERTIES, CTB WATER FOR WETTING				
CTB Water Used	Craig	CTB2	Craig	Craig
Moisture Level for Highest Strength Value, wt. percent	18	22	25	30
Highest 56-day Unconfined Compressive Strength Value, psi	122	41	167	207

The second sample listed, from a test using CTB2 (1800 ppm sodium, 5000 ppm sulfate) for dilution water, showed approximately 40 psi for an unconfined compressive strength (UCS) with either high quality or CTB water used for wetting. The fact that the high TDS water, CTB2, was used for dilution of the atomizer feed slurry and for temperature control water may have contributed to the poorer strength development in this sample. However, the quality of the water used to wet the waste samples did not appear to affect strength development.

It should be noted that even though the unconfined compressive strength for the second sample was below the 50 to 60 psi mentioned above, in most landfill situations the wastes are not truly unconfined, but instead are packed to a given level in a fill area that is confined on several sides. Thus, only the very edge of the landfill approaches being truly unconfined. It is very likely that a material exhibiting a 40 psi UCS would be quite acceptable for landfilling and would allow operating heavy equipment on that landfill.

Another laboratory measurement of relative waste disposal properties is the permeability coefficient for the cured sample. This coefficient is a measurement of the rate at which water moves through a column of the cured wastes. Values of 10^{-7} cm/sec or smaller (e.g., 10^{-8} cm/sec, 10^{-9} cm/sec) are considered low enough to qualify a material as a pond liner. Most wet FGD stabilized sludges exhibit values in the 10^{-4} to 10^{-6} cm/sec range.

The permeability coefficient values measured for the same four samples are summarized in Table 5. As above, values are included for samples that

TABLE 5. WASTE SAMPLE PERMEABILITY COEFFICIENT DATA

SAMPLE DESCRIPTION				
Sample No.	1	2	3	4
Inlet SO_2 Level, ppm	400	1000	1000	2000
Recycle Ratio	12:1	2.1:1	1.7:1	1.4:1
Reagent Ratio	0.9	1.1	1.6	1.3
Overall SO_2 Removal, %	92	98	98	92
Reagent Utilization, %	98	89	62	71
Dilution Water Used in FGD System	Service	CTB2	Service	Service
WASTE PROPERTIES, DEIONIZED WATER FOR WETTING				
Permeability Coefficient, cm/sec	7.8×10^{-6}	1.3×10^{-4}*	3.0×10^{-6}	1.7×10^{-6}
Standard Deviation	1.4×10^{-6}	0.8×10^{-4}	0.8×10^{-6}	0.5×10^{-6}
WASTE PROPERTIES, CTB WATER USED FOR WETTING				
CTB Water Used	Craig	CTB2	Craig	Craig
Permeability Coefficient, cm/sec	4.3×10^{-6}	2.4×10^{-5}	3.5×10^{-6}	6.1×10^{-6}
Standard Deviation	0.8×10^{-6}	0.8×10^{-5}	0.8×10^{-6}	1.5×10^{-6}

* Value determined in an experimental effort conducted two years earlier.

** Value not available.

were wetted with deionized water and cured and for samples that were wetted with CTB water and cured. Presented are mean values, which represent the arithmetic mean of 10 determinations, and the standard deviations about those means. The samples from the first, third, and fourth tests listed show permeability coefficients in the 10^{-6} cm/sec range. The use of the Craig CTB to wet these three samples prior to curing had no consistent effect on permeability coefficient. One value was decreased by a factor of two, one was nearly equal, and the third was three to four times greater than those for deionized water. Differences of less than one order of magnitude, as are these, can be considered insignificant.

In the second test listed, CTB2 (1800 ppm Na^+, 5000 ppm $SO_4^=$) was used for all system makeup except lime slaking water. The permeability coefficient values range from 10^{-4} to 10^{-5} cm/sec, with the sample wetted with CTB actually measuring at the more desirable lower value. The greater permeability coefficient values in the 10^{-4} to 10^{-5} cm/sec range for this sample are probably related to the poorer strength development observed and may be a result of the high sodium levels in the waste sample. The permeability coefficient value using deionized water rather than CTB2 water for wetting was measured at an earlier time, so differences in laboratory operator techniques and differences in exact curing conditions may account for a portion of the difference of nearly one order of magnitude between the two values. However, in spite of the observation that the use of CTB 2 as process makeup water may have had a negative impact on waste disposal properties, the use of CTB2 for wetting the waste sample prior to curing did not have a negative effect.

In summary, all of the waters appeared to have an acceptable impact on system SO_2 removal performance when used for all makeup other than lime slaking. In fact, high sodium or chloride levels may have beneficial effects on SO_2 removal. Sulfate concentration appears to be a good indicator as to whether the water can be used for slaking as well. However, other species such a sodium or chloride may offset any negative effects of sulfate ion. Tests with a water with a 700 ppm sulfate concentration showed no negative effects on slaked lime utilization in the pilot system, while tests with waters with a sulfate content of 2000 ppm and greater showed obvious negative effects. Tests with a water with a 1000 ppm sulfate content were inconclusive because of the beneficial effects of chloride ions present. Consequently, at a sulfate ion concentration between 700 ppm and 2000 ppm (in the absence of high sodium or chloride concentrations) the sulfate present begins to have significant negative impacts on slaked lime reactivity.

Given the limited number of measurements presented here, and the unknown nature of the correlation between laboratory measurements and actual landfill properties, it is difficult to relate waste disposal properties to the use of CTB waters in the FGD process. However, in all of the measurements made, the use of CTB water for wetting the solid waste material appeared to have no significant negative impacts on cured waste strength and permeability coefficient values. Future testing will measure impacts of the use of CTB waters on waste leachate quality.

CALCIUM CHLORIDE AS AN SO_2 REMOVAL PERFORMANCE ADDITIVE

Tests at Arapahoe with a high chloride content makeup water and results reported elsewhere [4,5,6] indicate that a substantial improvement in SO_2 removal and corresponding lime utilization can result from a high chloride content (e.g., 1.0 percent as Cl) in the solids produced in a spray dryer/fabric filter FGD system. Benefits in both spray dryer and fabric filter SO_2 removal at a given reagent ratio have been observed. The benefit of the chloride is thought to result from the deliquescent properties of calcium chloride. A deliquescent substance is one that has an affinity for moisture and that correspondingly dissolves in absorbed moisture. Such substances tend to be hard to dry, leading to increased droplet residence times in the spray dryer, and higher residual moisture levels in the solids collected on the fabric filter surfaces. Longer droplet residence times may result in increased time available for liquid phase SO_2 removal reactions, while higher residual moisture levels have been linked to increased SO_2 removal across fabric filter cakes[4].

Discussion of the results of the chloride addition tests is divided into benefits observed in system SO_2 removal, and other, potentially negative, impacts on the FGD system.

Impacts on System SO_2 Removal Performance

Most of the calcium chloride addition tests conducted at Arapahoe have been for higher inlet SO_2 levels (2000 ppm and greater), because it is at higher SO_2 levels where the cost benefits of improved lime utilization are greatest. Also, inlet SO_2 levels of 2000 ppm and greater will require SO_2 removals of 90 percent or greater under current New Source Performance Standards. This becomes increasingly more difficult to achieve in a spray dryer based FGD system as inlet SO_2 levels are substantially increased.

Figure 6 shows the results of chloride addition tests for a 2000 ppm inlet SO_2 level. Data for two chloride levels (0.4 wt. percent and 0.8 wt. percent chloride in the fabric filter solids collected) are represented as open and dark symbols, respectively. The smooth curves in Figure 6 represent a visual fit of SO_2 removal data for baseline tests (no chloride added) at a 2000 ppm inlet SO_2 level. All tests are at a 20°F approach, and recycle to achieve approximately a 30 wt. percent solids level in the atomizer feed slurry.

For the two tests at 0.4 wt. percent chloride in the fabric filter solids and a reagent ratio of around 1.1, overall SO_2 removal is improved by approximately 5 to 10 percentage points. The improvement appears to be evenly split between the spray dryer and fabric filter contributions to overall removal. At higher reagent ratio values of 1.3 to 1.4, similar improvements in overall removal of between 5 and 10 percentage points are observed. However, the benefits appear to lie entirely within the spray dryer at these reagent ratios. This observation is misleading, though, because overall removal levels of essentially 100 percent are observed. Even though the fabric filter contribution to overall SO_2 removal is not

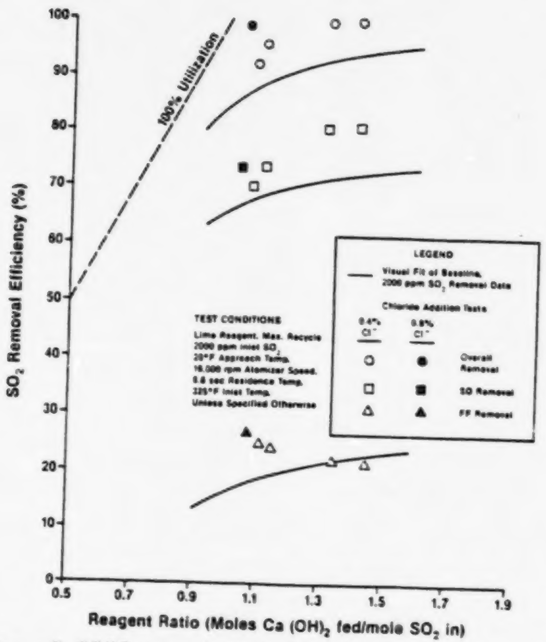


Figure 6. Results of 2000 ppm inlet SO_2 level tests with chloride addition compared to visual fit of 2000 ppm baseline data.

increased, the fabric filter is removing essentially 100 percent of the SO_2 reaching the bags.

Only one point is shown for a 0.8 wt. percent chloride test. In this test, nearly 100 percent SO_2 removal is achieved at a reagent ratio between 1.0 and 1.1. This amounts to nearly a 15 percent improvement in overall SO_2 removal relative to baseline (no chloride addition) performance at the same reagent ratio. This overall improvement is split between improvements in spray dryer and fabric filter contributions, and results in a lime utilization of over 90 percent.

These data indicate that SO_2 removal levels of greater than 90 percent can be achieved at a reagent ratio of approximately 1.1 with a relatively low chloride level of 0.4 percent, and approximately 1.0 at a higher chloride level of 0.8 wt. percent. Without chlorides, a reagent ratio of approximately 1.3 or better is required. The corresponding improvement in lime utilization amounts to an increase from approximately 72 percent without chlorides to approximately 85 percent at a 0.4 wt. percent chloride level in the fabric filter solids, and over 90 percent at a 0.8 wt. percent chloride level.

Because of the improvements in SO_2 removal capabilities and in lime utilization observed, tests were conducted at spray dryer inlet SO_2 levels as high as 3000 ppm. The results of 3000 ppm tests are presented in Figure 7. The curves in Figure 7 represent baseline (no chloride addition) SO_2 removal performance at a 2000 ppm inlet SO_2 level. The open symbols illustrate the decrease in SO_2 removal performance with an increase in inlet SO_2

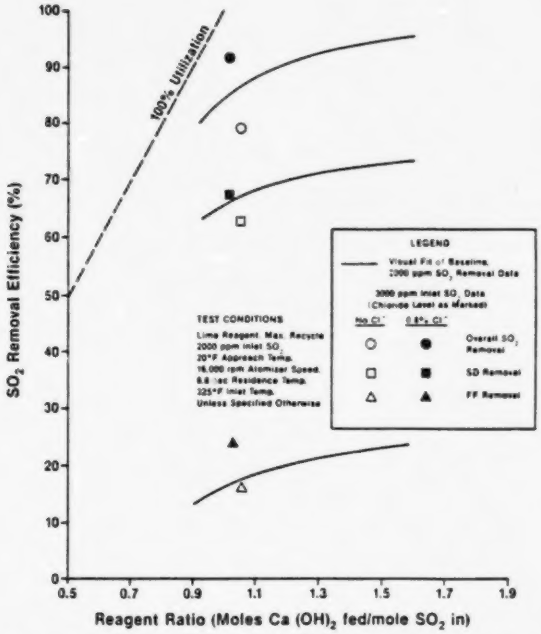


Figure 7. Results of 3000 ppm inlet SO_2 level tests with and without chloride addition compared to visual fit of 2000 ppm baseline data.

level from 2000 ppm to 3000 ppm, still with no chlorides added. The increase in inlet SO_2 level results in a reduced overall SO_2 removal by approximately 5 to 10 percentage points. It should be noted that increased inlet SO_2 level requires larger quantities of lime to achieve a given reagent ratio. For a constant solids level in the atomizer feed slurry, the ratio of recycle to fresh lime material is correspondingly reduced. Consequently, increased inlet SO_2 concentrations can have the effect of reducing the recycle ratio that is achievable at a given reagent ratio, due to limitations on the weight percent solids level that can practically be achieved. This reduction in recycle level may partly account for the reduced SO_2 removal in the higher inlet SO_2 tests.

One test was conducted at a 3000 ppm inlet SO_2 concentration with chloride added to the 0.8 wt. percent level. In this test, the overall SO_2 removal increased to just over 90 percent. This corresponds to an observed increase in lime utilization from approximately 75 percent without chlorides to 82 percent. Similar benefits were observed at a 2500 ppm inlet SO_2 level.

Thus, calcium chloride addition appears to result in improved lime utilization and/or increased SO_2 removal capabilities in high sulfur operation (inlet SO_2 levels of 2000 ppm or greater). However, some negative impacts on the operation and maintenance of the spray dryer/ fabric filter system may result from chloride addition. Some of these impacts are described below.

Other Impacts of Chloride Addition

Based on a limited number of observations, chloride addition has a potentially negative impact in three areas: fabric filter pressure drop, stress corrosion cracking of 300 series stainless steels in the system, and solid waste disposal characteristics.

The effects of chloride addition on fabric filter pressure drop were determined inadvertantly. One long-term goal of this pilot unit is to determine spray dryer effects on fabric filter performance. Fabric filter pressure drop values are closely monitored, particularly the increase in pressure drop with time after new bags are put into service. In previous papers it had been noted that the fabric filter pressure drop downstream of spray dryers appeared to be lower than for those in hot, untreated flue gas service[2,3]. However, no data were available to compare pressure drop performance for the two modes of operation under similar conditions of length of time in service and air-to-cloth ratio.

Recently, two compartments of the 10 MW fabric filter pilot plant at Arapahoe were rebagged at the same time. One compartment, downstream of the pilot spray dryer, is the subject of this paper. The other is a control compartment treating normal, 275°F flue gas. The control compartment was operated at an air-to-cloth ratio of 2:1. For most tests, the compartment downstream of the spray dryer was operated at the same air-to-cloth ratio. Based on previous observations, it was expected that the compartment downstream of the spray dryer would operate at an equal or slightly lower pressure drop. However, within several days of startup with new bags, the pressure drop of the compartment downstream of the spray dryer was significantly higher than that of the control compartment. This can be observed by comparing Figure 8, the pressure drop history of the control compartment, with Figure 9, that for the compartment downstream of the spray dryer.

Several of the dramatic increases seen in Figure 9 correspond to operating the compartment downstream of the spray dryer at atypical conditions, such as higher air-to-cloth ratios or longer times between compartment cleaning. However, even accounting for these periods, several step changes in compartment pressure drop are observed. These effects are illustrated in Figure 10, which shows the pressure drop data from Figure 9 replotted with data during atypical conditions removed. Also marked on Figure 10 are dashed lined that represent visual average pressure drops during various operating periods, and descriptors for spray dryer tests that correspond to step changes in these pressure drop averages.

This figure shows that on four occassions, noted by an increase in the level of the dashed line representing a visual average for the data, rapid increases in the tubesheet pressure drop occurred. In each case, the increase occurs immediately after a chloride addition test, or in one case a high sodium content CTB test. The high sodium test is presumed to have similar effects to that of the chloride tests, of producing moisture-retaining solids. It is not certain, though, what mechanisms contributed to the increased pressure drop values, or whether the increases are long-term or short-term in nature. Also, much of the spray dryer testing during this

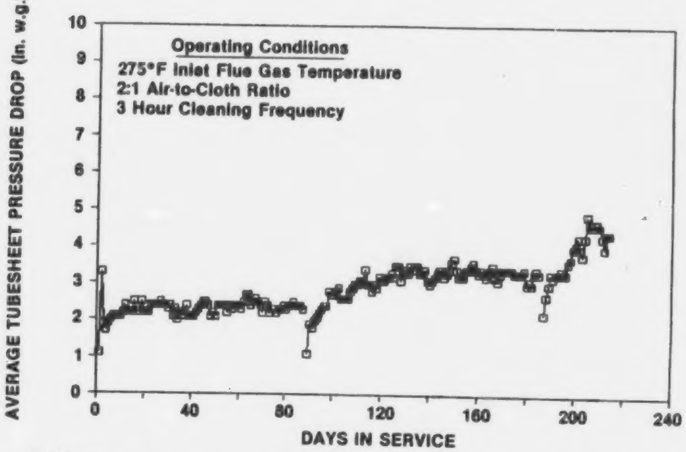


Figure 8. Tubesheet pressure drop values for fabric filter control compartment.

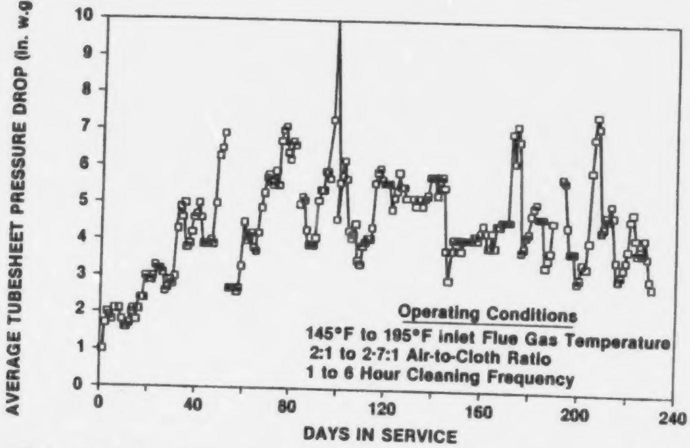


Figure 9. Tubesheet pressure drop values for fabric filter compartment downstream of spray dryer.

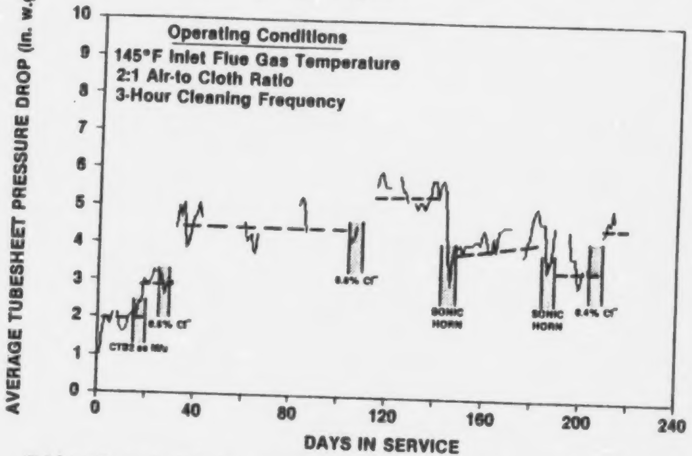


Figure 10. Edited tubesheet pressure drop values for fabric filter compartment downstream of spray dryer.

period resulted in fabric filter operation at varied conditions of air-to-cloth ratio, cleaning frequency, and inlet mass loading. Pressure drop data from periods of fabric filter operation at conditions that were not normal with respect to these values were editted from Figure 10. However, these atypical operating periods may have had continued effects on the remaining pressure drop data. It should be noted that sonic horn tests at approximately 140 to 180 days were sufficient to restore pressure drops to control compartment levels (compare pressure drop values for approximately 200+ days in Figures 8 and 10).

Although these observations as to the effects of chlorides on fabric filter pressure drop are not conclusive, they do indicate a potential for problems. Future long term chloride addition tests are planned to resolve these issues.

Chloride-related stress corrosion cracking was observed on 300 series stainless steel bag clamps during chloride addition tests. Seven of 36 clamps failed in a matter of days. Metallurgical failure analysis identified classic evidence of stress corrosion cracking, and high chloride levels were measured by micro-probe analyses near the crack surfaces[7]. While neither alarming nor particularly surprising, this documented failure indicates that 300 series stainless steels may not be suitable for slurry or gas contact (even on the "clean gas" side of the fabric filter) if chloride addition is contemplated.

The final apparent negative impact of chloride addition is on solid waste disposal properties. Table 6 summarizes measured unconfined compressive strength (UCS) values for wetted, compacted and cured samples from tests at three chloride levels: 0 (none added), 0.4 wt. percent, and 0.8 wt. percent in the fabric filter solids. The three tests are at similar levels of inlet SO_2 concentration (2000 ppm), overall SO_2 removal (90 percent or greater), and approach temperature (20°F). As the chloride level is increased, the sample UCS values decrease. In fact, the sample at the highest chloride level of 0.8 wt. percent in the fabric filter solids resulted in a maximum UCS of 28 psi. This value is marginal for a landfill on which heavy equipment is to be operated.

A confounding effect is that as the chloride level is increased, the amount of excess lime present is decreased (i.e., utilization increases). Lime is a known reactant in cementation reactions, so the lower excess lime levels may contribute to lower developed strength. However, samples from tests with no chloride addition during low sulfur operation (400 ppm to 1000 ppm inlet SO_2 levels) with nearly 100 percent lime utilization, or with virtually no excess lime, showed no tendencies for reduced strength. Future laboratory studies will involve lime addition to these high chloride test samples to determine if sample strengths are increased.

Table 7 summarizes permeability coefficient data for cured samples from these same three tests. As for the USC values, the permeability coefficient values become less desirable (increase) as the chloride content is increased. Again, future laboratory tests will determine if lime addition can improve (decrease) these values for compacted and cured samples.

TABLE 6. WASTE SAMPLE UNCONFINED COMPRESSIVE STRENGTH VALUES FOR VARYING CHLORIDE LEVELS

SAMPLE DESCRIPTION			
Sample No.	4	5	6
Inlet SO_2 Level, ppm	2000	2000	2000
Recycle Ratio	1.4:1	1.8:1	1.2:1
Reagent Ratio	1.3	1.1	1.1
Overall SO_2 Removal, %	92	90	99
Reagent Utilization, %	71	81	91
Chloride Level in Solids, wt. %	0.0	0.4	0.8
WASTE PROPERTIES, DEIONIZED WATER FOR WETTING			
Moisture Level for Highest Strength Value, wt. %	30	30	22
Highest 56-day Unconfined Compressive Strength, psi	293	168	28

TABLE 7. WASTE SAMPLE PERMEABILITY COEFFICIENT VALUES FOR VARYING CHLORIDE LEVELS

SAMPLE DESCRIPTION			
Sample No.	4	5	6
Inlet SO_2 Level, ppm	2000	2000	2000
Recycle Ratio	1.4:1	1.8:1	1.2:1
Reagent Ratio	1.3	1.1	1.1
Overall SO_2 Removal, %	92	90	99
Chloride Level in Solids, wt. %	0.0	0.4	0.8
WASTE PROPERTIES, DEIONIZED WATER FOR WETTING			
Permeability Coefficient, cm/sec	1.7×10^{-6}	5.4×10^{-6}	2.1×10^{-5}
Standard Deviation	0.5×10^{-6}	1.7×10^{-6}	0.4×10^{-5}

In summary, chloride addition to levels below 1.0 percent chloride in the waste solids can produce significant improvements in lime utilization and/or maximum SO_2 removal achievable. However, chloride addition may have negative impacts in the areas of fabric filter pressure drop performance, suitability of 300 series stainless steels for use in the system, and solid waste landfill properties. Future tests will further study these effects.

WARM-GAS-BYPASS REHEAT

For the purposes of this paper, warm-gas-bypass reheat is defined as a bypass of spray dryer inlet gas around the spray dryer in order to raise the temperature of the flue gas entering the fabric filter. In comparison, hot-gas-bypass reheat would accomplish the same effect by ducting a lesser amount of air heater inlet gas to the spray dryer outlet, although at a loss in boiler efficiency.

Reheat of the spray dryer outlet gas may be desirable as an approach to reducing corrosion tendencies on fabric filter compartment walls. Not only are fabric filter corrosion tendencies reduced, but the percentage of flue gas treated by the spray dryer is reduced by 5 to 30 percent, possibly allowing a smaller vessel to be used. Potentially negative impacts of warm gas bypass include the increased volume of flue gas to be treated in the fabric filter, possibly requiring a larger fabric filter, and the fact that overall SO_2 removal may be limited because a portion of the flue gas sees no SO_2 removal in the spray dryer. Also, lab-scale data indicate that decreased relative humidity in the fabric filter, one result of warm gas bypass, reduces SO_2 removal tendencies across the filter cake.[4]

A number of warm-gas-bypass reheat tests have been conducted on the Arapahoe pilot unit. Results of several tests conducted at a 1000 ppm inlet SO_2 level are summarized below.

Figure 11 compares 95 percent confidence intervals for baseline 1000 ppm, no reheat, 20°F approach temperature tests to results of tests which raised the fabric filter inlet flue gas temperature by 20°F. The open symbols in Figure 12 represent overall SO_2 removal values for tests in which warm gas bypass was used to raise the spray dryer outlet temperature by 20°F. This amounted to approximately a 15 percent bypass around the spray dryer. Overall removal levels are reduced by 5 to 10 percentage points. The overall reagent ratio (moles of lime added per mole of SO_2 in the total flow of flue gas to the system) required to achieve 70 percent overall SO_2 removal was increased from about 0.75 without reheat to 0.9 to 1.0 with 20°F of reheat. An overall SO_2 removal level of 80 percent could be achieved at a 0.9 to 1.0 reagent ratio in baseline tests, but required a 1.3 reagent ratio with 20°F of warm-gas-bypass reheat. In terms of reagent utilization, 70 percent SO_2 removal was achieved at a lime utilization of over 90 percent in baseline tests, but only 70 to 80 percent with 20°F of reheat. Reagent utilization at an 80 percent overall SO_2 removal level dropped from approximately 85 percent to less than 65 percent.

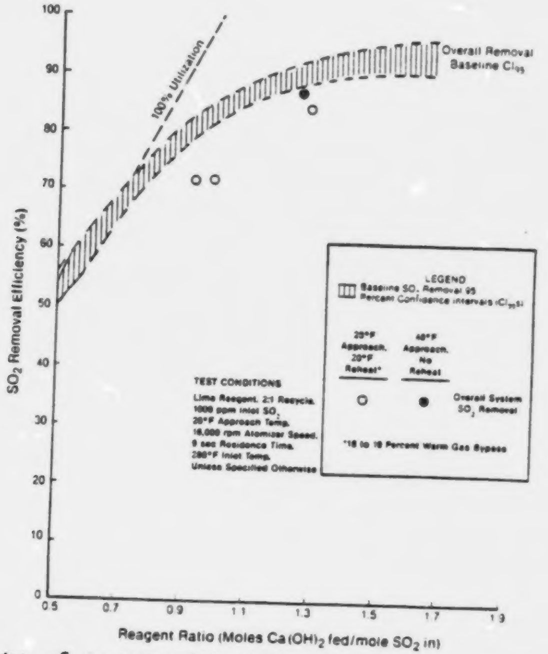


Figure 11. Results of tests with 20°F of warm-gas-bypass reheat and a test at a 40°F approach temperature compared to baseline test 95 percent confidence intervals.

The dark symbol in Figure 11 represents the results of a single, 40°F approach temperature test with no gas bypass. This approach temperature results in the same flue gas temperature at the fabric filter inlet as for the 20°F reheat, 20°F approach temperature tests, but in this case all of the gas goes through the spray dryer. These results indicate that 40°F approach temperature operation, treating all of the flue gas in the spray dryer, has less of an effect on SO_2 removal and lime utilization. However, as this is only a single test result, further tests will be required to determine the tradeoffs between 20°F of warm gas bypass with a 20°F approach versus operation at a 40°F approach.

Figure 12 compares the results of 40°F warm-gas-bypass reheat tests to 95 percent confidence intervals for baseline, 1000 ppm, 20°F approach temperature tests. These results show major decreases in overall removal levels achieved. In fact, the data lead to doubts that overall SO_2 removal levels of 70 percent or greater could be achieved. Warm gas reheat to these levels has two negative effects on SO_2 removal: 1) nearly 30 percent of the total flue gas treated by the system bypasses the spray dryer, thus limiting overall SO_2 removal, and 2) the relative humidity of the flue gas at the fabric filter inlet is greatly reduced. The latter effect has been observed to have negative effects on fabric filter SO_2 removal capabilities[4].

These results indicate that for systems treating flue gas with a 1000 ppm inlet SO_2 level, warm gas bypass may not be the most desirable method of avoiding fabric filter compartment wall corrosion tendencies. This appears to be particularly true for systems that require high SO_2 removal

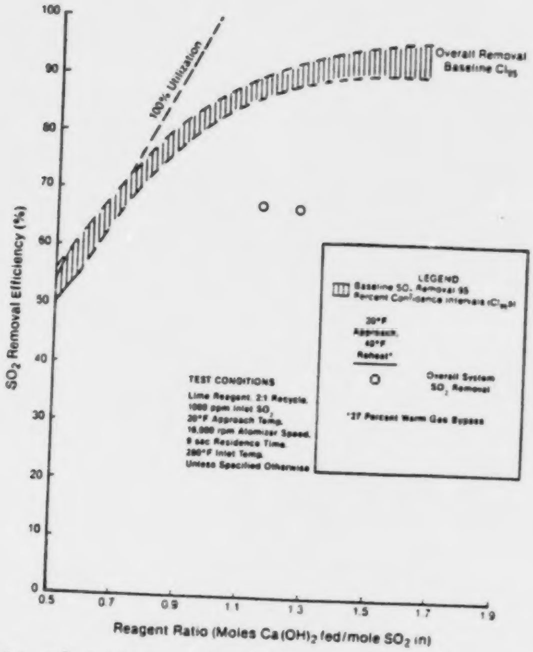


Figure 12. Results of tests with 40°F of warm-gas-bypass reheat compared to baseline test 95 percent confidence intervals.

levels (above 80 percent). Future testing will determine if smaller quantities of hot gas reheat (air heater inlet gas bypass) can achieve similar fabric filter protection at a lesser impact on SO_2 removal.

UPDATE OF FABRIC FILTER CORROSION TENDENCIES

Previous papers[2,3] have noted that an alarming tendency has been observed for corrosion of pilot fabric filter compartment walls. The corrosion appeared to be related to low compartment wall temperatures during low ambient air temperature operation. This was largely attributed to inadequate insulation of the compartment walls, because the fabric filter was originally intended to treat hot (275°F) flue gas, and to large proportional heat losses resulting from the relatively high surface area to volume ratio of a pilot-scale compartment.

After the corrosion was noted, several measures were implemented:

- Compartment wall temperatures were monitored on a continuous basis,
- The compartment was sandblasted and coated, and
- Steps were taken to reduce heat losses and cold air infiltration.

The following is a brief discussion of the results of these efforts.

Compartment wall temperatures were found to drop as low as 96°F during 9°F ambient temperature operation. This resulted in certain condensation of flue gas moisture, as the flue gas dewpoint was approximately 104°F.

Coatings were seen as a preventative measure to ensure the integrity of the fabric filter compartment walls for the remaining one to three years of spray dryer pilot plant testing. A relatively inexpensive vinyl ester coating (Cook 931-B-301) was applied to the compartment walls. The compartment doors were covered with test patches of the vinyl ester, and two (more expensive) epoxy coatings: ConChem Fibercrete AR and Duromar HPL3100. After over one year of operation, the vinyl ester coating has been quite effective in preventing the dramatic exfoliating corrosion originally observed. Minor patches of corrosion are seen in cold spots and/or in the vicinity of "holidays" in application of the coating, but the condition of the walls is greatly improved.

The epoxies on the doors, which due to air inleakage are in a more severe corrosion service, have proven very effective. A sandblasted, but uncoated patch on one door exhibits the severe exfoliation of corrosion byproducts expected. A vinyl ester coated area is corroded, but to a much lesser extent. The two areas coated with epoxies show slight discoloration, but essentially no corrosion. These results indicate that spot coating of full-scale fabric filter compartment walls can be very effective in eliminating or reducing corrosion near "cold spots." In minor corrosion areas, less expensive vinyl ester coatings may be sufficient, but in severe service areas, more expensive epoxies may be required.

A final update on fabric filter corrosion tendencies is to report the effects of improved insulation and improved air inleakage control on compartment wall temperatures. Improvements in these areas on the Arapahoe pilot fabric filter compartment have been limited to insulating a previously uninsulated shaker box (not currently used) on one compartment wall, and replacing door seals. Table 8 summarizes minimum wall temperature data in two low ambient temperature cases. Case 1 was measured soon after fabric filter wall corrosion tendencies were observed. A wall temperature as low as 96°F was measured during a period of 9°F ambient temperature. Case 2 at the same ambient temperature was measured after the improvements listed above. The minimum wall temperature measured was over 20°F above that measured previously, and appeared to be above the flue gas dewpoint. This dramatic improvement in minimum wall temperature measured illustrates that a major factor in avoiding fabric filter compartment wall corrosion will be attention to detail in the areas of insulation installation and maintenance, and prevention of ambient air inleakage.

TABLE 8. SUMMARY OF AMBIENT TEMPERATURE EFFECTS ON PILOT-SCALE FABRIC FILTER COMPARTMENT WALL TEMPERATURES

	Before Improved Insulation, Replaced Door Seals	After Improved Insulation, Replaced Door Seals
Ambient Temperature, °F	9	9
Spray Dryer Outlet Temperature, °F	137	137
Flue Gas Dewpoint, °F	104	104
Minimum Wall Temperature Measured, °F	96	117

CONCLUSIONS

The testing discussed in this paper evaluates several modifications to the basic spray dryer/fabric filter dry FGD process. Conclusions about each of these potential modifications are summarized below.

- Use of Cooling Tower Blowdown Water - Testing with four simulated and two actual CTB waters indicated that all were acceptable for use as atomizer feed slurry dilution water and/or temperature control water. The acceptability of these waters for use in lime slaking appeared to be related to sulfate concentration in the water. Laboratory tests with two of the waters showed them to have little or no measurable negative impact when used to wet waste solids before compacting and curing. These test results indicate that CTB waters may be largely substituted for fresh waters for use in a spray dryer-based dry FGD system. However, testing should be conducted before implementing the use of higher sulfate content (over 700 ppm) waters for lime slaking.

- Use of Calcium Chloride as a Performance Additive - These results indicate that calcium chloride addition to achieve chloride levels of up to one percent in the fabric filter solids can be quite effective in promoting high SO_2 removal levels and high lime reagent utilization. These benefits appear to be particularly desirable for systems in high sulfur operation, at inlet SO_2 levels of 2000 ppm or above. Chlorides may be useful to extend the applicability of spray dryer/fabric filter FGD technology to higher inlet SO_2 systems. However, several potential negative impacts of chloride addition have been noted. These include increases in fabric filter pressure drop related to chloride addition, chloride-induced stress corrosion cracking failure of 300 series stainless steels in flue gas service, and apparent decreases in solid waste disposal properties as chloride levels are increased. The potential problems should be further investigated before chloride addition is implemented on a full scale system.

- Warm-Gas-Bypass Reheat to Prevent Fabric Filter Corrosion - These results indicate that for low sulfur systems, with inlet SO_2 levels of 1000 ppm or less and requiring overall SO_2 removal levels of 80 percent or less, warm gas bypass to provide up to 20°F of reheat may be effective if the decreased lime utilization is acceptable. To provide higher fabric filter inlet temperatures or for higher inlet SO_2 levels and higher SO_2 removal requirements, the penalty in lime utilization may be such that other approaches such as hot gas bypass are more cost effective.

- Effectiveness of Fabric Filter Corrosion Prevention Measures - Coatings in the pilot fabric filter compartment have been largely effective in reducing tendencies for corrosion on wall surfaces. An inexpensive vinyl ester coating has provided good protection, while two more expensive epoxies appear to be providing excellent protection. Also, preventive measures such as insulation of small, previously uninsulated surfaces and replacement of compartment door seals have been effective in raising minimum wall temperatures observed during low ambient temperature operation. These results demonstrate that corrosion in fabric filters downstream of a spray dryer operated at 20°F approach temperature conditions can be substantially reduced or avoided by improved insulation practices, reducing ambient air infiltration, and spot coating of problem areas.

REFERENCES

1. Blythe, Gary M. et al, "EPRI Spray Dryer Pilot Plant Status and Results," presented at the joint EPA/EPRI Symposium on Flue Gas Desulfurization, Hollywood, Florida, May 17-20, 1982.

2. Blythe, Gary M., and Richard G. Rhudy. "EPRI Spray Dryer/Baghouse Pilot Plant Status and Results," presented at the EPA/EPRISymposium on Flue Gas Desulfurization, New Orleans, Louisiana, November 1-4, 1983.

3. Rhudy, Richard G., and Gary M. Blythe, "Fabric Filter Operation Downstream of a Spray Dryer: Pilot and Full-Scale Results." Presented at the EPA/EPRI Joint Particulate Control Symposium, Kansas City, Kansas, August 27-30, 1984.

4. Karlsson, Hans T., et al. "Activated Wet-Dry Scrubbing of SO_2," Journal of the Air Pollution Control Association, Volume 33, No. 1, January 1983 pp. 23-28.

5. Hansen, Svend Keis, et al, "Status of the Joy/Niro Dry FGD System and Its Future Application for the Removal of High Sulfur, High Chloride and NO_x from Flue Gas." Publication No. 83-JPGC-APC-8, The American Society of Mechanical Engineers, New York, NY.

6. Blythe, Gary M., and Richard G. Rhudy, "Field Evaluation of a Utility Dry FGD System." Presented at the EPA/EPRI Symposium on Flue Gas Desulfurization, New Orleans, Louisiana, November 1-4, 1983.

7. Fuentes, Karen. Memorandum to Gary Blythe, Radian Corporation, Austin, Texas, July 30, 1984.

PERFORMANCE OF A SPRAY DRYER/ESP FLUE GAS CLEANUP SYSTEM DURING TESTING AT THE PITTSBURGH ENERGY TECHNOLOGY CENTER

James T. Yeh, Charles J. Drummond, and James I. Joubert
United States Department of Energy
Pittsburgh Energy Technology Center
Pittsburgh, PA 15236

Dennis M. Tulenko, Robert R. Crynack, and Martin L. Hochhauser
Wheelabrator Air Pollution Control
Pittsburgh, PA 15219

ABSTRACT

Wheelabrator Air Pollution Control and the United States Department of Energy jointly sponsored a project to evaluate the potential for the application of spray dryer flue gas desulfurization (FGD) systems to existing coal-fired boilers equipped with electrostatic precipitators for particulate emissions control. The objectives of this project were to characterize the performance of a spray dryer FGD system using flue gas from the combustion of different types of coal, and to study the effectiveness of an electrostatic precipitator (ESP) in controlling particulate emissions from this system. Three different coals (low-, medium-, and high-sulfur coal) were burned in a 500 lb/hr pulverized-coal combustion test facility at the Pittsburgh Energy Technology Center. Sulfur dioxide was removed from the flue gas by injection of a lime slurry in a spray dryer. Particulate emissions were controlled using a mobile electrostatic precipitator. For each fuel, a three-part test program was conducted. This test program consisted of (1) base-line ESP performance tests without spray dryer operation to determine requirements for fly ash collection; (2) parametric tests defining sulfur dioxide removal efficiency and ESP collection efficiency at various spray dryer operating conditions; and (3) sorbent recycle tests in which the particulate collected by the ESP was recycled by injection into the spray dryer with the fresh sorbent to improve reagent utilization. Test results show that sulfur dioxide removal efficiencies of 90 percent can be achieved for each of the three coals tested using the combined spray dryer and electrostatic precipitator system for emissions control. Performance criteria for the electrostatic precipitator were met during all phases of testing.

INTRODUCTION

With increasing concern over sulfur dioxide emissions, many electric power generating utilities are evaluating the modifications that would be required to equip existing power plants with flue gas desulfurization systems. One of the flue gas desulfurization processes being considered is the installation of a spray dryer before the existing particulate collector. In most existing coal-fired electric power generating facilities, the particulate collection device is a high-efficiency electrostatic precipitator. A test program was conducted between September, 1983, and June, 1984, to evaluate the potential for the application of spray dryer flue gas desulfurization systems to existing coal-fired boilers equipped with electrostatic precipitators for particulate emissions control.

The use of dry-scrubbing technology for the control of sulfur dioxide emissions from coal-fired electric power generating facilities is attractive to utility and other industries because of the ease of operation and the anticipated economic advantages it can offer compared to wet scrubbing [1]. Considerable progress has been made in the past several years in adapting conventional spray-drying technology for the control of sulfur dioxide emissions from coal-fired industrial and utility boilers [2-15]. A spray dryer flue gas desulfurization (FGD) system uses an alkaline sorbent material injected into the flue gas as an aqueous solution or slurry; the sorbent reacts with the sulfur dioxide (SO_2) to form a dry product containing sulfates and sulfites [16-17]. A mixture of spent sorbent and fly ash is removed from the flue gas stream by an electrostatic precipitator (ESP) or by a fabric filter.

Dry FGD systems have predominantly been installed on new facilities with fabric filters used for particulate control. As a result, few investigations of spray dryer/ESP systems have been conducted, and limited information and operating experience is available, for the application of spray dryers to existing facilities using electrostatic precipitators for particulate control.

In this study, the combined performance of a spray dryer and an electrostatic precipitator was evaluated. Three coals were used, a 1.6 percent sulfur Pittsburgh seam coal, a 2.7 percent sulfur Pittsburgh seam coal, and a 0.7 percent sulfur Kentucky Harlem seam coal. Analyses of the coals used in these tests are given in Table 1. For each type of coal fired, a three-part test program was conducted. The first part evaluated ESP performance without the spray dryer in operation (base-line tests). The second part involved operation of the spray dryer at 70, 80, and 90 percent overall sulfur dioxide removal efficiencies while measuring ESP performance. The third part of the program evaluated ESP performance during the injection of a fly ash and spent sorbent mixture (recycle tests) that had been collected from the previous spray dryer tests conducted with fresh slurry.

TABLE 1. Typical Analyses of Coals Used

	Coal I Pittsburgh Seam	Coal II Pittsburgh Seam	Coal III Kentucky
PROXIMATE ANALYSIS WT.% (Dry Basis)			
Volatile Matter	37.5	39.6	38.9
Fixed Carbon	53.0	52.1	57.3
Ash	9.5	8.3	3.8
ULTIMATE ANALYSIS WT.% (Dry Basis)			
Hydrogen	5.3	5.3	5.8
Carbon	74.0	73.5	78.0
Nitrogen	1.2	1.0	1.7
Sulfur	1.6	2.7	0.7
Oxygen	8.4	9.2	10.0
Ash	9.5	8.3	3.8
HEATING VALUE (Btu/lb)	13,314	13,490	14,240
MOISTURE IN COAL (As Fired)	1.0	1.1	1.0
ASH ANALYSIS			
Al_2O_3	22.68	21.68	26.47
CaO	1.30	1.19	2.07
Fe_2O_3	12.76	23.58	10.52
K_2O	1.99	1.46	1.70
Na_2O	0.30	0.36	0.41
SiO_2	58.37	47.37	51.12
MgO	0.84	0.94	1.60
SO_3	1.22	1.63	2.00

DESCRIPTION OF THE TEST FACILITY

Combustion Test Facility

The 500 lb/hr (227 kg/hr) combustion test facility is shown schematically in Figure 1. The furnace was designed to simulate the performance of an industrial or utility steam generator. The furnace walls are refractory-lined and water-cooled. The unit is 7 ft (2.13 m) wide, 5 ft (1.52 m) deep, and 12 ft (3.66 m) high, and has a volumetric heat liberation rate of about 16,000 Btu/hr-ft^3 (165,550 J/sec-m^3) at a thermal input of 6.5 million Btu/hr (6.86 (10^9) J/hr). The flue gas flow rate is approximately 1300 scfm.

Coal is charged to the hopper, pulverized to a size consist of 70 percent minus-200 mesh, and then conveyed by the primary air into a recycle coal loop where intimate mixing of coal and air occurs. Four adjustable exit tubes are connected to the recycle loop; these convey the primary air-coal mixture to each of the four burners. Secondary air at 600°F (316°C) is fed through adjustable swirl vanes surrounding each burner. Approximately 20 percent excess air is used. The flue gas exits the furnace at about 2000°F (1093°C), passes through a convective heat transfer section, and is then used to preheat the secondary air to the desired inlet temperature. By controlling the air through the recuperative air preheater, the flue gas exit temperature can be maintained in the range of 300°-475°F (149°-246°C).

Spray Dryer and Electrostatic Precipitator

The spray dryer and electrostatic precipitator used in these studies are shown in Figure 2. Sorbent slurries were prepared in separate facilities and transferred to the slurry storage tank. The sorbent used in this study was freshly slaked lime prepared on site by mixing 1 part of quicklime to 2.8 parts water. The exothermic reaction produced a temperature rise of between 110°F (43.3°C) and 120°F (49°C) over a 3-4 minute period. The composition of the quicklime used is given in Table 2. A centrifugal circulating pump and aeration were both used to maintain a uniform suspension of solids in the slurry. A progressing cavity pump, equipped with a variable-speed drive, provided the desired slurry flow rate. The slurry flow was measured by a mass flowmeter. The slurry was diluted with water near its point of entry to the spray dryer in order to maintain the desired flue gas temperature at the exit of the spray dryer.

The 7-ft (12.3-m)-diameter spray dryer and all ancillary equipment used in this study are identical to a unit used in a three-year pilot effort conducted by Wheelabrator Air Pollution Control at their Joliet, Illinois, pilot facility. The spray dryer is equipped with a 6-in. (15.24-cm)-diameter disc atomizer containing three 1/4-in. (0.635-cm)-diameter holes. The disc can be rotated at speeds up to 23,000 rpm, corresponding to a maximum tip speed of 600 ft/sec (183 m/sec). The vertical wall of the spray dryer is 4 ft (1.22 m) high, ending in a 60° bottom cone that tapers down to a 12-in. (30.5-cm)-diameter flue gas line. This flue gas line at the bottom of the spray dryer carries the stream of flue gas and the mixture of fly ash and spent sorbent to the electrostatic precipitator.

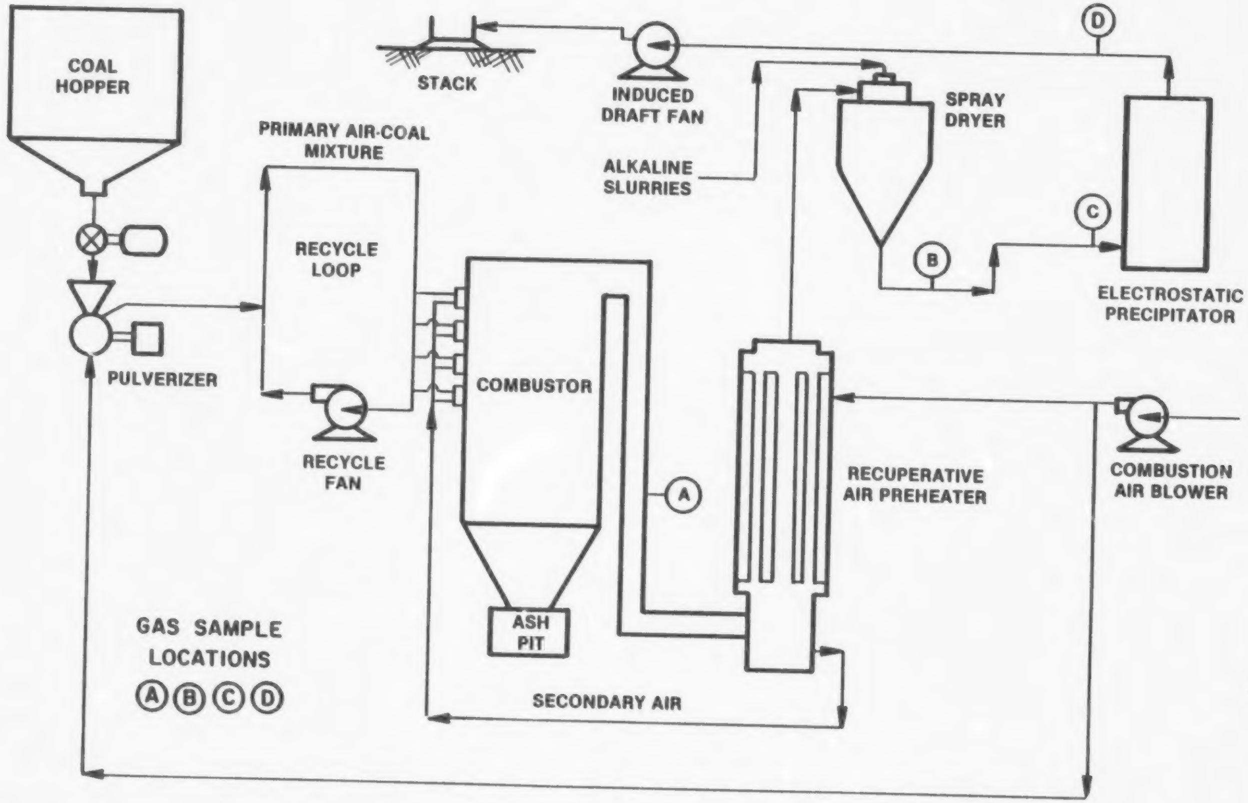

COAL HOPPER
STACK
INDUCED DRAFT FAN
SPRAY DRYER
PRIMARY AIR-COAL MIXTURE
ALKALINE SLURRIES
D
C
B
RECYCLE LOOP
PULVERIZER
RECYCLE FAN
COMBUSTOR
A
RECUPERATIVE AIR PREHEATER
ELECTROSTATIC PRECIPITATOR
COMBUSTION AIR BLOWER
ASH PIT
GAS SAMPLE LOCATIONS
ⒶⒷⒸⒹ
SECONDARY AIR

FIGURE 1. SIMPLIFIED FLOW SHEET OF 500 LB/HR PULVERIZED-COAL-FIRED TEST FACILITY.

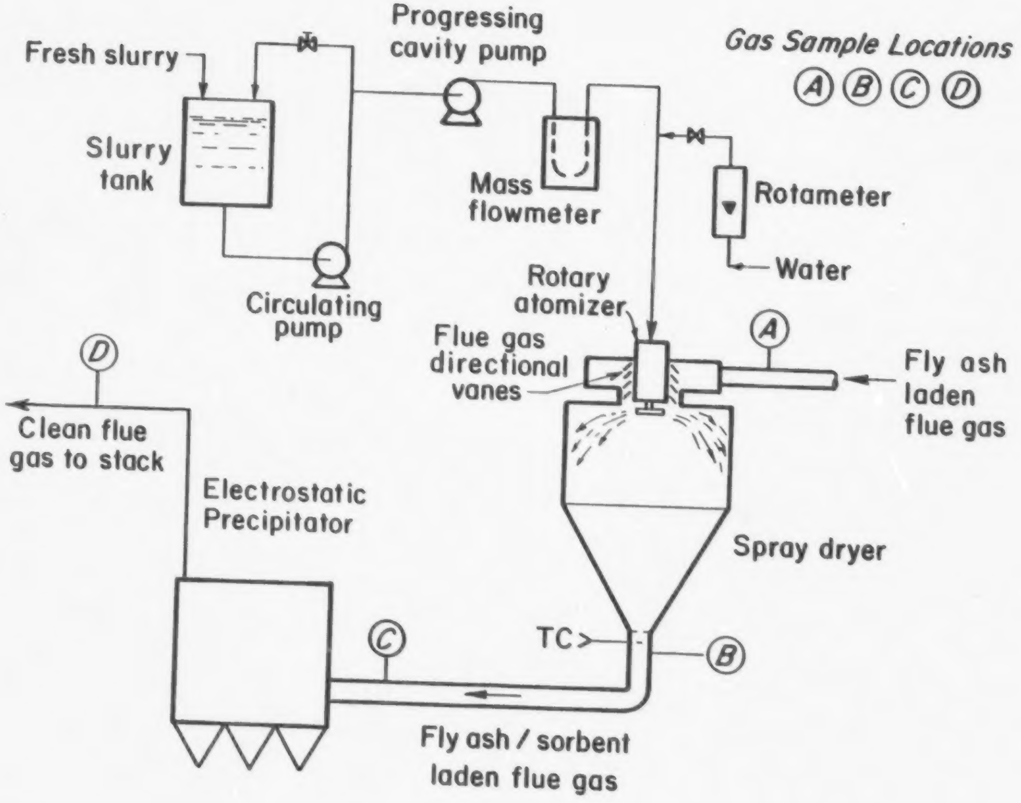

Gas Sample Locations
Ⓐ Ⓑ Ⓒ Ⓓ
Fresh slurry
Slurry tank
Progressing cavity pump
Mass flowmeter
Rotameter
Water
Circulating pump
Rotary atomizer
Flue gas directional vanes
Ⓐ
Fly ash laden flue gas
Ⓓ
Clean flue gas to stack
Electrostatic Precipitator
Spray dryer
TC
Ⓑ
Ⓒ
Fly ash / sorbent laden flue gas

FIGURE 2. SPRAY DRYER/PRECIPITATOR SO_2 REMOVAL SYSTEM.

TABLE 2. Chemical Analysis of Quicklime

	Percent
Calcium Oxide - Available	94.00
Calcium Carbonate	2.00
Silica	1.20
Aluminum Oxide	0.32
Ferric Oxide	0.13
Magnesium Oxide	0.60
Sulfur	0.025
Calcium Sulfate	0.055
Phosphorus Pentoxide	0.013
Manganese Oxide	0.003

Flue gas enters the spray dryer at about 400°F (204°C) and passes through a set of vanes that surround a rotary atomizer. These vanes, as shown in Figure 3, are designed to provide adequate mixing of the flue gas with the alkaline slurry being injected through the rotary atomizer. The dried sorbent and fly ash are then carried by the flue gas to the electrostatic precipitator, where the particulates are removed and the clean flue gas is discharged to the stack. The plug flow gas residence time in the spray dryer is estimated to be 7-8 seconds (based on spray dryer exit temperature).

The electrostatic precipitator used in these studies was a Wheelabrator Air Pollution Control mobile electrostatic precipitator (MESP) [18]. This electrostatic precipitator is a three-field unit complete with automatic voltage controls, transformer/rectifier sets, rappers, vibrators, hoppers, and an induced draft fan capable of delivering 9000 acfm (254.9 actual cubic meters per minute) at 700°F (371°C). The mobile electrostatic precipitator is mounted on a trailer 8.0 ft wide, 54.5 ft long, and 13.5 ft high (2.44 m wide, 16.61 m long, and 4.12 m high).

Each of the three fields consists of sets of discharge electrodes and collecting plates. The standard discharge electrodes consist of pipe frames that are 5 ft high and 6.25 ft long (1.52 m high and 1.91 m long), having 12 wires per frame. The Wheelabrator standard star-shaped wire was used in this program. The collecting plate is 6.5 ft (1.98 m) high and consists of 4 interlocking strips, providing a length of 6.25 ft (1.91 m) for each field.

Each collecting plate is individually rapped at the bottom with a rotating, falling hammer. All of the discharge electrodes in each field are cleaned with side-mounted vibrators. Both the pilot rappers and the discharge electrode vibrators are manually controlled for each field.

The ash- or dust-removal system consists of two trough hoppers with screw conveyors per field. The screw conveyors discharge into two pyramidal hoppers that empty into 55 gallon drums through a vacuum system.

Data Collection System

Flue gas constituent concentration, temperature, pressure, and flow were recorded using a computerized data collection system. Flue gas was sampled and analyzed at four locations (see Figure 1): the furnace outlet; the spray dryer outlet; the precipitator inlet; and the precipitator outlet.

The dew point of the flue gas was calculated using a computer code developed at the Pittsburgh Energy Technology Center [19]. The code requires information on the fuel composition, the fuel feed rate, the amount of excess air, the sorbent slurry concentration and feed rate, the dilution water feed rate, the moisture content of the combustion air, and the spray dryer system pressure, to compute the dew point, the flue gas composition, the moles of flue gas produced, and the air-to-fuel ratio. The code automatically corrects the calculated dew point for system pressure.

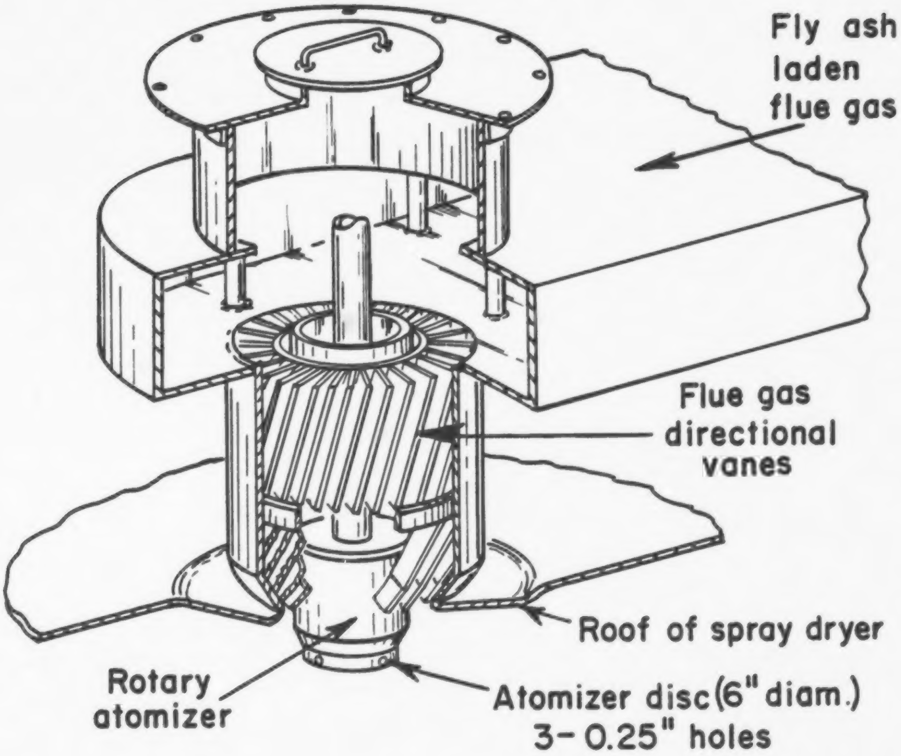

Fly ash laden flue gas
Flue gas directional vanes
Roof of spray dryer
Rotary atomizer
Atomizer disc (6" diam.) 3- 0.25" holes

FIGURE 3. SPRAY DRYER INLET ASSEMBLY.

RESULTS AND DISCUSSION

Sulfur Dioxide Removal

The sulfur dioxide removal efficiencies obtained in tests conducted without recycle are shown in Figure 4. The spray dryer reactor was operated at an average approach of 25F° (13.9C°) to the flue gas saturation temperature at the exit of the reactor. The approach-to-saturation temperature is the difference between the flue gas temperature at the exit of the spray dryer and the flue gas dew point at the exit of the spray dryer.

A 90 percent sulfur dioxide removal efficiency for the overall system (spray dryer reactor and electrostatic precipitator) was obtained at a Ca/S molar ratio of 2.8 (the Ca/S molar ratio is defined as the moles of available calcium in the sorbent slurry per mole of sulfur in the inlet flue gas).

Tests conducted using recycle of the fly ash collected in the ESP showed that this mode of operation was not effective in improving sorbent utilization or sulfur dioxide removal efficiency. Little available calcium was detected in the recycled fly ash.

Contribution of the Electrostatic Precipitator to Sulfur Dioxide Removal

The sulfur dioxide removal efficiency of the ESP is defined as follows:

$$\left[\frac{SO_2\ \text{entering ESP} - SO_2\ \text{leaving ESP}}{SO_2\ \text{entering ESP}}\right](100).$$

For example, if the spray dryer removes 80 percent of the sulfur dioxide entering the spray dryer, and the ESP removes 20 percent of the sulfur dioxide entering the ESP, then the total system sulfur dioxide removal efficiency will be 84 percent. Sulfur dioxide removal in the ESP occurs when the flue gas contacts unreacted sorbent collected on the plates of the ESP. The sulfur dioxide removal efficiency observed within the ESP was less than 20 percent on average for all conditions tested (based on the sulfur dioxide concentration entering the ESP). Because of the enhanced contact between unreacted sorbent and the flue gas that occurs in a fabric filter, the sulfur dioxide removal efficiency for fabric filters has been found to be about double that observed in an ESP at the same approach-to-saturation temperature [20]. However, industrial and utility boiler operators are willing to operate spray dryer/ESP systems at 10F° closer to the flue gas dew point than similar spray dryer/fabric filter systems [20]. This is because ESP systems are not susceptible to "blinding" by moist particulate matter when the flue gas approaches its dew point as are fabric filters. The lower operating temperature that can be maintained in spray dryer/ESP systems increases the sulfur dioxide removal efficiency obtained in the spray dryer, compensating for the less efficient sulfur dioxide removal in the ESP.

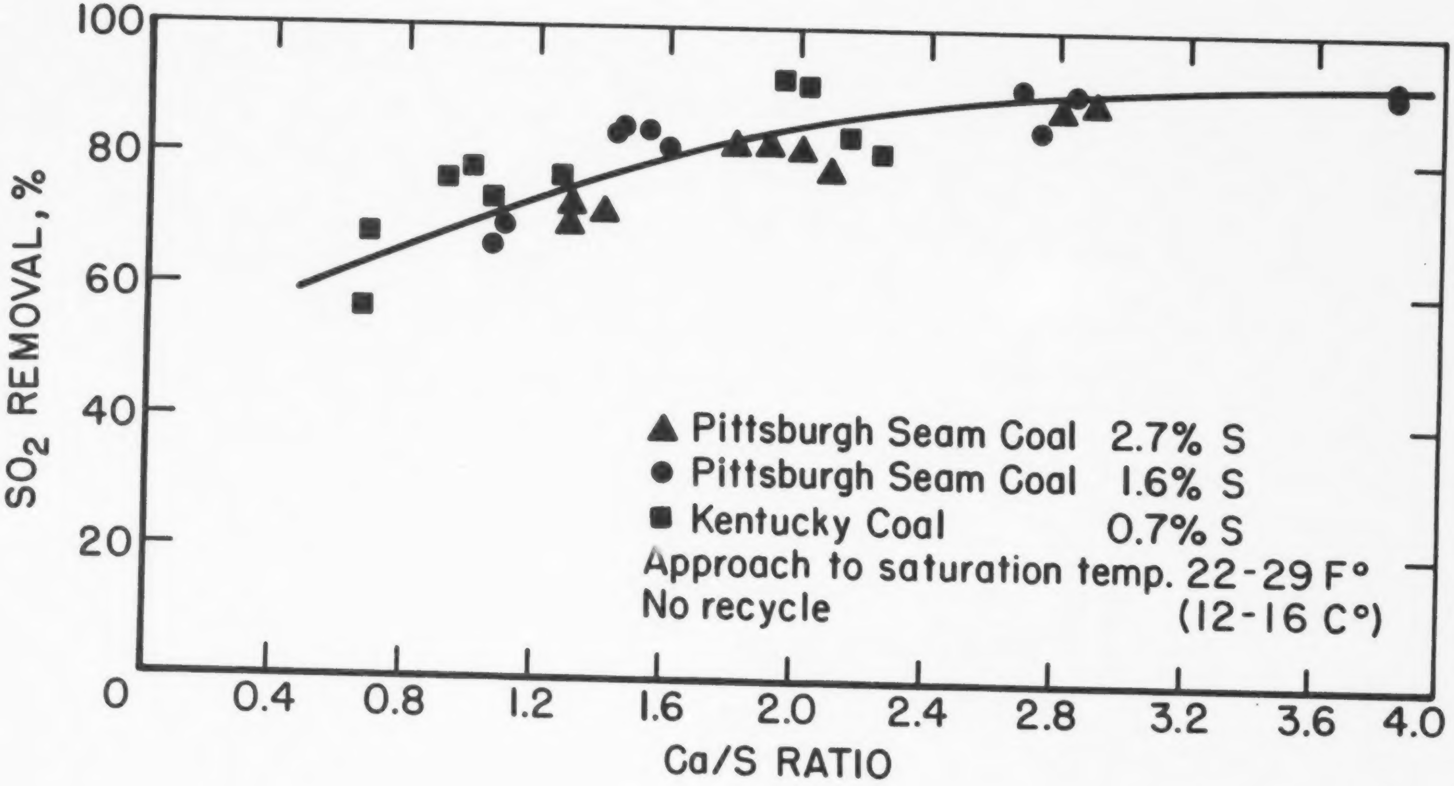


FIGURE 4. TOTAL SO$_2$ REMOVAL BY THE SPRAY DRYER/ESP SYSTEM.

Particulate Removal

The pilot ESP was operated at constant conditions during the entire test program to maintain consistent ESP operation, while spray dryer operating parameters were varied. The ESP operating conditions kept constant include the volumetric gas flow rate, the collecting surface area (i.e., specific collection area, or SCA, in ft^2/1000 acfm), and the power input (watts/1000 acfm). The performance of the ESP during each test was determined by the measured ESP particulate collection efficiency. The performance of electrostatic precipitators has traditionally been described by the Deutsch-Anderson Equation [21]:

$$E = 1-e^{-(A/V)w} \tag{1}$$

where
E = ESP particulate collection efficiency (%)
A = ESP collecting surface area (ft^2)
V = Volumetric flow rate of gas through ESP (ft^3/sec)
w = Migration velocity (ft/sec)

Since gas flow and collecting surface area were held constant during these tests, the measured ESP particulate collection efficiency is directly related to the migration velocity (w). Thus by comparing ESP particulate collection efficiencies, we are also comparing migration velocities.

A summary of the test results for the three coals studied is shown in Tables 3 and 4. A detailed discussion of the particulate removals obtained for each coal studied is given below.

ESP Performance While Firing Coal No. I

Coal No. I was a 1.6 percent sulfur Pittsburgh seam coal. The average ESP inlet particulate loading measured while the spray dryer was not operating (base-line tests) was 3.71 lb/10^6 Btu. Under these conditions, the ESP outlet particulate loading varied between 0.022 and 0.044 lb/10^6 Btu. An average ESP particulate collection efficiency of 99.03 percent was obtained during the base-line tests.

With the spray dryer operating at 70, 80, and 90 percent overall sulfur dioxide removal efficiencies, the inlet particulate loading increased with the slurry feed rate to 6.75, 7.20, and 12.31 lb/10^6 Btu, respectively. This increased particulate loading resulted in increased outlet residuals but also showed a trend toward increased ESP particulate collection efficiency over the base-line data. However, at a sulfur dioxide removal efficiency of 80 percent, the ESP particulate collection efficiency was lower than that observed during base-line testing. During tests conducted with sorbent recycle, the ESP particulate collection efficiency was slightly higher than at base-line tests but was lower than at 70% and 90% SO_2 removal levels.

TABLE 3. ESP Performance (No Sorbent Recycle)

	Coal I	Coal II	Coal III
Sulfur(%)	1.6	2.7	0.7
Ash (%)	9.5	8.3	3.8
BASE-LINE TESTS (SPRAY DRYER NOT OPERATING)			
Particulate Loading ($lb/10^6$ Btu)			
ESP Inlet	3.711	2.900	1.838
ESP Outlet	0.031	0.021	0.011
ESP Collection Efficiency (%)	99.03	99.23	99.41
SPRAY DRYER TESTS (NO SORBENT RECYCLE)[1]			
70% Overall SO_2 Removal Efficiency			
Ca/S Ratio[2]	1.08	1.33	0.86
SO_2 Removal (%)	68.6	71.8	71.7
Slurry Solids (lb/hr)	14.4	33.0	5.3
ESP Inlet Loading ($lb/10^6$ Btu)	6.748	12.160	3.373
ESP Outlet Loading ($lb/10^6$ Btu)	0.035	0.111	0.022
ESP Collection Efficiency (%)	99.48	99.08	99.34
80% Overall SO_2 Removal Efficiency			
Ca/S Ratio[2]	1.51	1.94	1.52
SO_2 Removal (%)	83.9	80.7	79.5
Slurry Solids (lb/hr)	21.1	45.3	9.6
ESP Inlet Loading ($lb/10^6$ Btu)	7.200	14.259	3.749
ESP Outlet Loading ($lb/10^6$ Btu)	0.074	0.115	0.040
ESP Collection Efficiency (%)	98.93	99.18	98.92
90% Overall SO_2 Removal Efficiency			
Ca/S Ratio[2]	3.18	2.82	1.98
SO_2 Removal (%)	90.0	88.3	92.5
Slurry Solids (lb/hr)	43.5	66.0	12.0
ESP Inlet Loading ($lb/10^6$ Btu)	12.306	17.630	4.800
ESP Outlet Loading ($lb/10^6$ Btu)	0.058	0.233	0.032
ESP Collection Efficiency (%)	99.54	98.68	99.30

Notes:

[1] Approach-to-saturation temperature at spray dryer outlet: 22F° to 29F° (12C° to 16C°).

[2] Moles of available calcium in the sorbent slurry per mole of sulfur in the flue gas.

lb/hr = 0.454 kg/hr

$lb/10^6$ Btu = 0.43 $kg/10^9 J$

TABLE 4. ESP Performance (With Sorbent Recycle)[1]

	Coal I	Coal II	Coal III
2:1 RATIO (LB FLY ASH/LB FRESH LIME)			
Ca/S Ratio[2]	2.02	1.18	0.72
Overall SO_2 Removal Efficiency (%)	82.2	69.5	55.1
Slurry Solids (lb/hr)	86.4	84.0	17.1
ESP Inlet Particulate Loading (lb/10^6 Btu)	15.510	18.965	4.468
ESP Outlet Particulate Loading (lb/10^6 Btu)	0.137	0.275	0.032
ESP Collection Efficiency (%)	99.12	98.55	99.30
4:1 RATIO (LB FLY ASH/LB FRESH LIME)			
Ca/S Ratio[2]	1.54	1.03	2.23
Overall SO_2 Removal Efficiency (%)	77.3	70.9	81.6
Slurry Solids (lb/hr)	84.4	85.0	42.0
ESP Inlet Particulate Loading (lb/10^6 Btu)	16.370	15.717	7.000
ESP Outlet Particulate Loading (lb/10^6 Btu)	0.140	0.182	0.086
ESP Collection Efficiency (%)	99.14	98.84	98.77

Notes:

[1] Approach-to-saturation temperature at spray dryer outlet: 22F° to 29F° (12C° to 16C°).

[2] Moles of available calcium in the sorbent slurry per mole of sulfur in the flue gas.

lb/hr = 0.454 kg/hr

lb/10^6 Btu = 0.43 kg/10^9J

ESP Performance While Firing Coal No. II

Coal No. II had a higher sulfur and lower ash content than Coal No. I. During base-line tests with this coal, average ESP inlet particulate loadings of 2.90 lb/10^6 Btu, average ESP outlet particulate loadings of 0.021 lb/10^6 Btu, and average ESP particulate collection efficiencies of 99.23 percent were observed.

Because of the higher sulfur content of Coal No. II, the spray dryer slurry feed rate had to be increased to obtain the same overall sulfur dioxide removal efficiency. The highest ESP particulate loading (17.63/10^6 Btu) occurred during spray dryer operation at 90 percent sulfur dioxide removal. Under these conditions, the ESP outlet residual was 0.233 lb/10^6 Btu, and the ESP particulate collection efficiency was 98.7 percent.

The trend observed during testing with Coal No. II was that of decreasing ESP particulate collection efficiency with increasing sulfur dioxide removal efficiency (increasing particulate loading). During tests conducted with sorbent recycle, the ESP particulate collection efficiency followed a similar trend, decreasing with increasing particulate loadings.

ESP Performance While Firing Coal No. III

Coal No. III was a low-sulfur, low-ash coal from Kentucky. Base-line tests with this coal resulted in average ESP inlet particulate loadings of 1.8 lb/10^6 Btu, the lowest of the three coals tested. The average ESP outlet particulate loading was 0.011 lb/10^6 Btu, resulting in an ESP particulate collection efficiency of 99.41 percent during base-line tests, the best performance observed while firing these three coal types.

Because of the low-sulfur content of Coal No. III, the spray dryer slurry feed rate required to obtain 70, 80, and 90 percent sulfur dioxide removal efficiencies was lower than for Coals No. I and II. As a result, the average ESP inlet particulate loadings ranged from 3.373 to 4.800 lb/10^6 Btu, while the average ESP outlet particulate loadings ranged from 0.022 to 0.040 lb/10^6 Btu. The ESP particulate collection efficiencies ranged from 98.9 percent to 99.3 percent.

When fly ash and spent sorbent collected during initial spray dryer tests were recycled to the spray dryer, the average ESP inlet particulate loadings increased, ranging from 4.5 to 7.0 lb/10^6 Btu. The average ESP outlet particulate loadings were observed to be 0.03 and 0.09 lb/10^6 Btu at recycle ratios of 2:1 and 4:1, respectively. The ESP particulate collection efficiencies for these recycle tests were 99.3 and 98.8 percent, respectively.

The particulate collection trend for Coal No. III was similar to that for Coal No. II. The ESP particulate collection efficiencies decreased from the base-line ESP particulate collection efficiency when the sorbent slurry was injected in the spray dryer. The observed particulate collection efficiencies for tests using sorbent recycle were inconclusive, as the particulate collection efficiencies for a 2:1 recycle ratio are similar to the base line, while they are substantially lower than the base line for a 4:1 recycle ratio. This may be due to the higher ESP inlet particulate loadings in the 4:1 recycle ratio tests compared with the 2:1 recycle ratio tests.

Compliance with Emissions Control Regulations

The capacity of an existing ESP is important in determining whether an electric power generating facility will meet particulate emissions control standards after installation of sulfur dioxide emissions control equipment (such as a spray dryer). Based on the data previously reviewed in this paper, if an ESP is operating with outlet particulate loadings close to compliance levels, it may not meet emissions control standards after the addition of a spray dryer to the system unless modifications to the ESP, such as increasing power, improving plate alignment, improving gas distribution, or replacing aged components, are made. Many operating electrostatic precipitators were designed for "worst case" conditions, such as low-sulfur, high-resistivity-ash Western coal, but operate with easier to precipitate Eastern coal. The results from this study show that these units should have enough collection area to meet compliance with the higher particulate loading inherent in the operation of a spray dryer. Operating parameters, such as migration velocity and specific collecting area of the ESP, would have to be analyzed for each installation to determine whether the particulate collection efficiency of the ESP would be sufficient while operating a spray dryer.

CONCLUSIONS

The efficient removal of sulfur dioxide from coal combustion flue gas can be achieved with low-, medium-, and high-sulfur Eastern U.S. coals using a system combining a spray dryer and an electrostatic precipitator. The results of this investigation provide the information required to determine whether an existing electric power generating facility can install a spray dryer for sulfur dioxide emissions control while using an existing ESP for particulate emissions control. The data show that the major parameter affecting ESP performance was the inlet particulate loading, which increased with increasing sorbent slurry flow rates. As a result, the ESP outlet residuals increased. However, the ESP particulate collection efficiency was about the same whether or not the spray dryer was operating except when the inlet particulate loadings were very high (either during operation to achieve 90 percent sulfur dioxide removal efficiencies or with sorbent recycle.) In addition to achieving good particulate collection efficiencies in the ESP, handling of the collected mixture of fly ash and spent sorbent posed no problems.

In determining whether a spray dryer can be used for sulfur dioxide emissions control at an electric power generating facility equipped with an ESP, the operating parameters of the existing ESP must first be evaluated. If particulate emissions just meet emissions standards, the ESP will most probably not meet particulate emissions standards after installation of a spray dryer without modifications. If the ESP easily meets emissions standards, the system will probably also meet emissions standards while operating a spray dryer. It can generally be concluded that the behavior of an electrostatic precipitator receiving flue gas treated in a spray dryer is similar to that normally expected for a conventional cold-side electrostatic precipitator in a coal-fired application.

Acknowledgment

The assistance of John D. Sherow of Wheelabrator Air Pollution Control, Precipitator Research and Development, in the ESP data analysis for this study was most helpful and greatly appreciated.

REFERENCES

1. Burnett, T.A.; Anderson, K.D.; and Torstrick, R.L. Spray Dryer FGD: Technical and Economic Assessment, Symposium on Flue Gas Desulfurization, Houston, Texas, EPA-600/9-81-0196, October 28-31, 1980, pp. 713-729.

2. Lutz, S.J.; Christman, R.C.; McCoy, B.C.; Mulligan, S.W.; and Slimak, K.M. Evaluation of Dry Sorbents and Fabric Filtration for FGD, prepared by TRW, Inc., for the U.S. EPA, EPA-600/7-79-005, January 1979.

3. Eriksen, P.E. The Development of Dry Flue Gas Desulfurization at Basin Electric Power Cooperative, Second Conference on Air Quality Management in the Electric Power Industry, Austin, Texas, January 22-25, 1980.

4. Sulfur Oxides Control Technology Series: Flue Gas Desulfurization-Spray Dryer Process, U.S. EPA Summary Report, EPA-625/8-82-009, September, 1982.

5. Parsons, E.L., Jr.; Kragh, O.T.; Brna, T.G. SO_2 Removal by Dry FGD, EPA Symposium on Flue Gas Desulfurization, Houston, Texas, October 28-31, 1980.

6. Downs, W.; Sanders, W.J.; Miller, C.E. Control of SO_2 Emissions by Dry Scrubbing, American Power Conference, Chicago, Ill., April 21-23, 1980.

7. Hurst, T.B.; Bielawski, G.T. Dry Scrubber Demonstration Plant Operating Results, EPA Symposium on Flue Gas Desulfurization, Houston, Texas, October 28-31, 1980.

8. Felsvang, K. Results from Operation of Riverside Dry Scrubber, the Riverside Dry FGD Symposium, Minneapolis, Minnesota, June 1981.

9. Farrington, J.F., Jr.; Lillestolen, T. Flakt's Dry FGD Technology, Flakt Engineering Technical Bulletin, Vol. 4, No. 4, December, 1980.

10. Adreasen, J. Flue Gas Desulfurization by Spray Dryer Absorption, Vol. 3, Energy and Environment Division, 11th World Energy Conference, Munich, West Germany, September 8-12, 1980.

11. Blythe, G.M.; Burke, J.M.; Kelly, M.E.; Rohlack, L.A.; Rhudy, R.G. EPRI Spray Drying Pilot Plant Status and Results, EPA FGD Symposium, Hollywood, Fla., May, 1982.

12. Henry, J.M.; Robards, R.F. Spray Dryer Scrubbers for High-Sulfur Coal Combustion, American Institute of Chemical Engineers National Meeting, New Orleans, La., November 8-12, 1981.

13. Yeh, J.T.; Demski, R.J.; Gyorke, D.F.; Joubert, J.I. Experimental Evaluation of Spray Dryer Flue Gas Desulfurization for Use with Eastern U.S. Coals, EPA FGD Symposium, Hollywood, Fla., May, 1982.

14. Farber, P.S. Start-Up and Performance of a High Sulfur Dry Scrubber System, Air Pollution Control Association Annual Meeting, New Orleans, La., June 20-25, 1982.

15. Sawyers, L.E.; Smith, P.V.; Caravano, C.; Jankura, B.J. Eastern Coal Spray Dryer Evaluation, Final Report to U.S. Department of Energy by Babcock and Wilcox, DE-AC22-81FE-17056, November, 1984.

16. Boynton, R.S. Chemistry and Technology of Lime and Limestone. John Wiley & Sons, Inc., New York, 2nd Ed., 1980.

17. Masters, K. Spray Drying Handbook, John Wiley & Sons, Inc., New York, 1979.

18. Crynack, Robert R.; Sherow, J.D. Use of a Mobile Electrostatic Precipitator for Pilot Studies, the Fifth Symposium on the Transfer and Utilization of Particulate Control Technology, Technical paper TP84-115, August, 1984.

19. Yeh, J.T.; Gyorke, D.F.; Pennline, H.W.; Drummond, C.J. Microcomputer Aided Calculations of Parameters for Spray Dryer Operation, Air Pollution Control Association Annual Meeting, Detroit, Mich., June 16-21, 1985.

20. Wilkinson, J.M.; Tonn, D.P. Baghouse vs. Precipitator for Dry Scrubber Systems -- Pilot Study Results, Coal Technology Conference, Houston, Texas, November 17-19, 1981.

21. Katz, J. The Art of Electrostatic Precipitation, S&S Printing Company, Inc., Pittsburgh, Pa., 1979.

DISCLAIMER

This report was prepared as an account of work sponsored by an agency of the United States Government. Neither the United States Government nor any agency thereof, nor any of their employees, makes any warranty, express or implied, or assumes any legal liability or responsibility for the accuracy, completeness, or usefulness of any information, apparatus, product, or process disclosed, or represents that its use would not infringe privately owned rights. Reference herein to any specific commercial product, process, or service by trade name, trade mark, manufacturer, or otherwise, does not necessarily constitute or imply its endorsement, recommendation, or favoring by the United States Government or any agency thereof. The views and opinions of authors expressed herein do not necessarily state or reflect those of the United States Government or any agency thereof.

APPLICATION OF ELECTROSTATIC ENHANCEMENT OF FABRIC FILTRATION TO SPRAY DRYER BYPRODUCT

by: L. S. Hovis and
Bobby E. Daniel
U.S. EPA/AEERL
Research Triangle Park, N.C. 27711
Yang-Jen Chen
Joy Industrial Equipment Co.
Los Angeles, CA 90039
R. P. Donovan
Research Triangle Institute
P.O. Box 12194
Research Triangle Park, N.C. 27709

ABSTRACT

Experiments conducted at small pilot scale show that the pressure drop increase during the fabric filtration of redispersed spray dryer byproduct (chiefly calcium salts and fly ash) is significantly reduced through electrostatic enhancement of the filtration. The pressure drop rise for a typical electrostatic augmented fabric filtration (ESFF) is only one-fourth or less than that of the rise for a conventional filtration cycle. The ESFF takes advantage of the relatively lower electrical resistivity of the spray dryer byproduct attributable to the high moisture content and to the low temperature at which the filtration of spray dryer byproduct is carried out. A low resistivity allows application of greater electrode potentials than can be used at normal fly ash filtration conditions.

This paper has been reviewed in accordance with the U.S. Environmental Protection Agency's peer and administrative review policies and approved for presentation and publication.

INTRODUCTION

A new, experimental baghouse facility at EPA's Air and Energy Engineering Research Laboratory in Research Triangle Park, N.C., was placed in operation in 1983 to fulfill the need for versatile laboratory baghouses which could duplicate field conditions with respect to temperature and at the same time accommodate innovations, especially electrostatic augmentation. This paper is primarily concerned with some of the initial results from experiments carried out in these baghouses, but also includes a brief description of the equipment and its operation.

This laboratory test facility consists of two one-bag compartments, one cleaned by reverse air and the other by pulse-jet action. These two compartments were built with independent controls which means that experiments can be carried out in each compartment at the same time. One of the first assignments for the reverse-air compartment was to perform experiments to answer the question related to an ESFF-temperature dependence which had been raised by reviewers at a 1983 peer review of ESFF. Since those initial experiments, the facilities have been used to test different fabrics, different dusts, precharging, electrode configurations, and other fabric filter and ESFF parameters. This paper reports only on the temperature dependence of fly ash ESFF and the application of ESFF to spray dryer byproduct solids.

DESCRIPTION OF EQUIPMENT

Each compartment of the baghouse test facility is based on the model used by Lamb and coworkers at Textile Research Institute (1), but has been designed to include high temperature operation and allow easy modification for nonstandard operating configurations. Each compartment can hold one bag up to 4 feet (1.2 m) long and about 8 inches (20 cm) in diameter. The compartments operate off one blower which delivers 100 cubic feet of air/minute (2.8 m^3/min). Dust is injected into the inlet air stream just before it enters the baghouse. Two separate feed hoppers are maintained, one for each compartment. A bypass system on the blower permits control of the gas flow independently to each compartment. A schematic diagram of the baghouse is shown in Figure 1. As shown in the figure, space for prechargers has been provided for, and, in fact, both corona wire prechargers and Masuda's Boxer Chargers (2) have been tried at these locations. Heating is provided by strip heaters on the baghouse proper, including the doors, and by pipe heaters clamped around the recirculating piping system. The whole system is well insulated except for the blower which is covered by a wooden box to cut down noise and to provide for some minimal thermal insulation. Figures 2 and 3 show the outside of the baghouse and the piping system, respectively. The blower cover and the control panel for the baghouse can be seen in Figure 3. Near atmospheric pressure is maintained by venting part of the system air to the room. So far this has worked well, but obviously the venting system would have to be changed if future experiments involved noxious gas addition to the air stream. Other than particulate matter, only water or steam has been injected into the air up to this time. Flow regulation and temperature control are automatic. The dust feeding system has not been automated, but this has not presented a major problem.

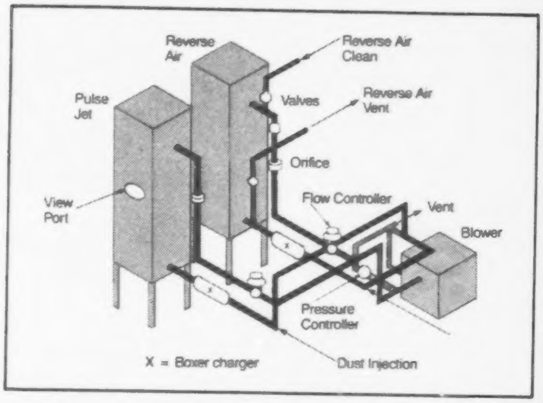

Reverse Air
Reverse Air Clean
Pulse Jet
Valves
Reverse Air Vent
Orifice
View Port
Flow Controller
Vent
Blower
x
x
Pressure Controller
X = Boxer charger
Dust Injection

Figure 1. Sketch of EPA high temperature experimental baghouse.

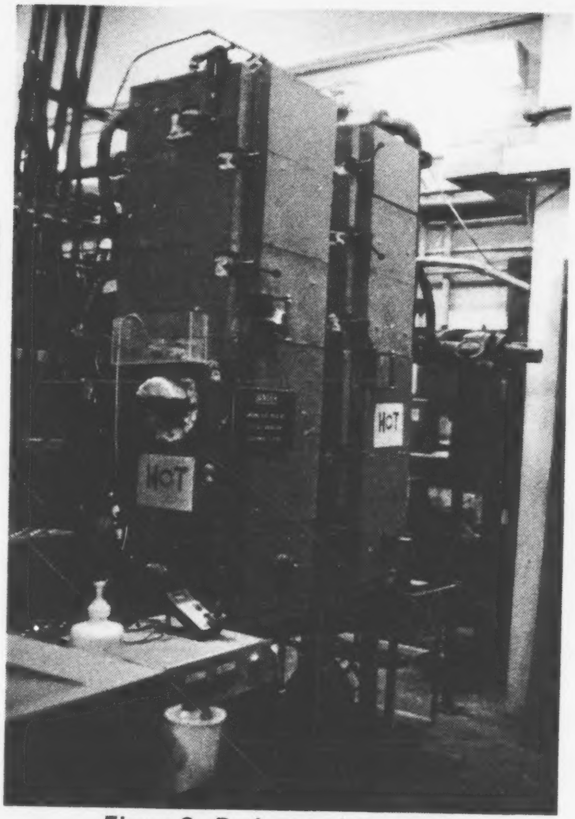


Figure 2. Baghouse photograph.

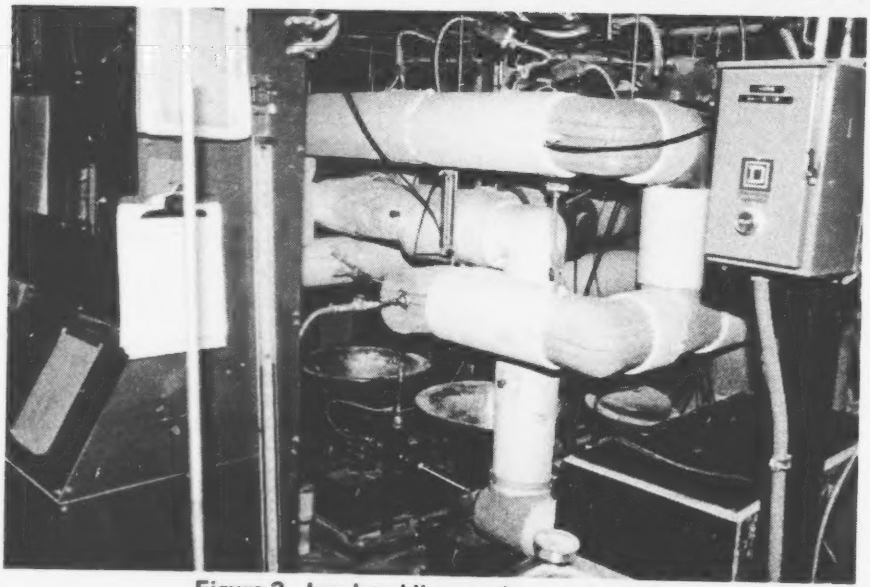


Figure 3. Insulated lines and control panel.

Filtration cycling and cleaning pulses are controlled by a Xanadu card controller.

The inside of the reverse-air compartment is shown in Figure 4. The unit is electrically connected for ESFF in the figure. Obviously both compartments were designed so that they could be electrified for ESFF. Also, glass covered sight ports with wipers were installed to observe the bag and/or dust during filtration.

EXPERIMENTS WITH FLY ASH

With completion of the pilot plant work on ESFF at the Du Pont Waynesboro, Virginia, site (3) and the installation of ESFF in the pilot plant at the Southwestern Public Services (SPS) Harrington Station using full length woven electrode bags (4), the in-house facilities described above became the primary laboratory scale unit for further experimental ESFF. One of the first experiments involved evaluation of ESFF on SPS dust at temperatures below 400° F (205° C), the standard test temperature at the Harrington Station pilot plant. The question of the temperature dependence of ESFF had arisen earlier at a peer review of ESFF. The high temperatures used at Harrington were judged to be greater than the operating temperatures of the majority of utility baghouses, and it was thought that results from Harrington could be biased because of this deviation. The reviewers recommended a supplementary laboratory study of temperature effects.

The initial study was a factorial experimental design involving temperature at two levels, gas velocities at two levels, and electrode field voltage at three levels (including zero voltage). The results of the experiment were predictable as far as velocity and electric field were concerned. Using the net pressure drop, ΔP_D, ($\Delta P_D = \Delta P_{final} - \Delta P_{eff}$) across the filter as the primary response, both gas velocity (A/C) and electric field had significant effects. The latter is the ESFF effect, in which ΔP_D decreases with increasing electric field. Increasing the gas velocity will increase ΔP_D in a given time cycle, but an interesting point here is that there is an interaction between the two variables in that the relative increase in ΔP_D that is brought about by higher A/C is reduced by the presence of an externally applied electric field. The plot of ΔP_D versus A/C in Figure 5 shows this interaction. However, the important new conclusion in this experiment is that a change in temperature over the range studied (300° to 350° F) (150° to 177° C) did not affect the ESFF performance. This conclusion is illustrated in Figure 6.

A second factorial test was carried out to examine the wider temperature range of 250° to 375° F (120° to 190° C). (The reverse-air in-house baghouse has a limit of about 375° F (190° C) because of heat losses through the fan.) The length of the filtration cycle and gas velocity were also varied as part of this second factorial test. The results were much the same. Figure 7 is a plot of the average ΔP_D across all the parameters versus temperatures in the range of 250° to 375° F (120° to 190° C) with power on (2 kV/cm field) and power off. The plot shows a lack of dependence of ΔP_D on temperature and no interaction of temperature and field. The conclusion reached by these experiments with fly ash is that high tempera-

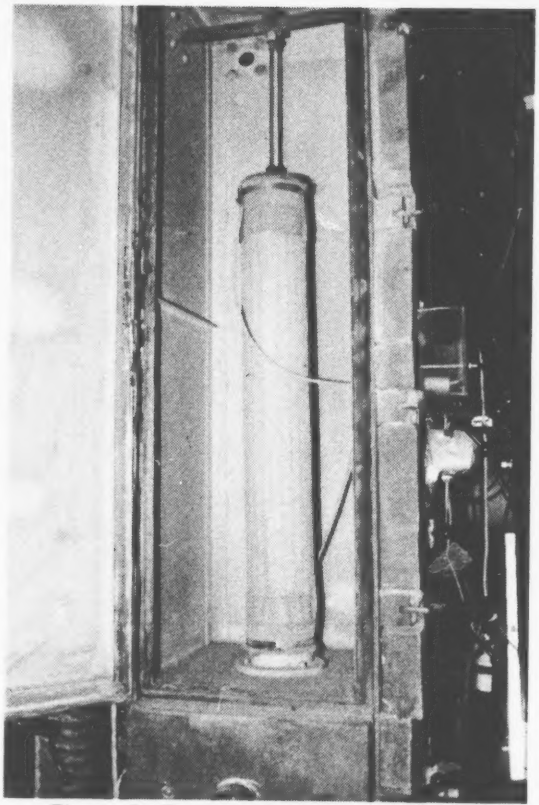


Figure 4. Reverse air bag mounting.

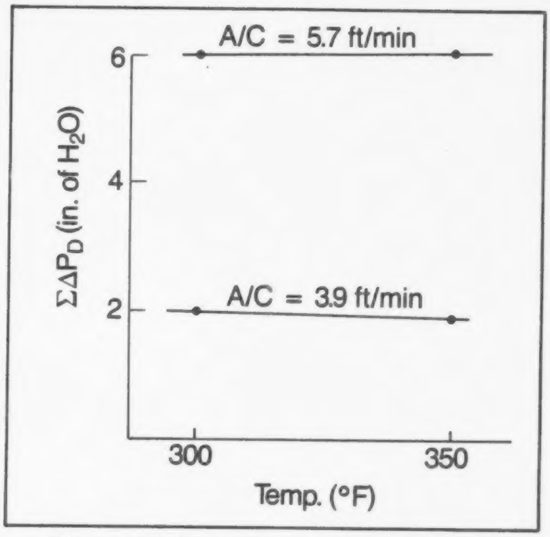


Figure 6. Temperature independence of electrical enhancement (reverse air, SPS fly ash).

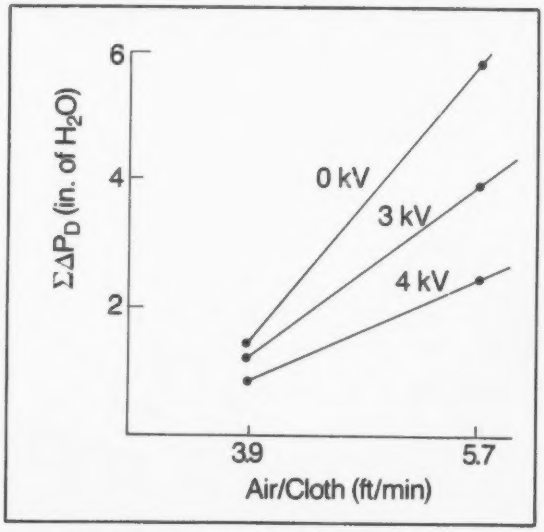


Figure 5. Net pressure drop at various face velocities and applied electrode voltages (reverse air, SPS fly ash).

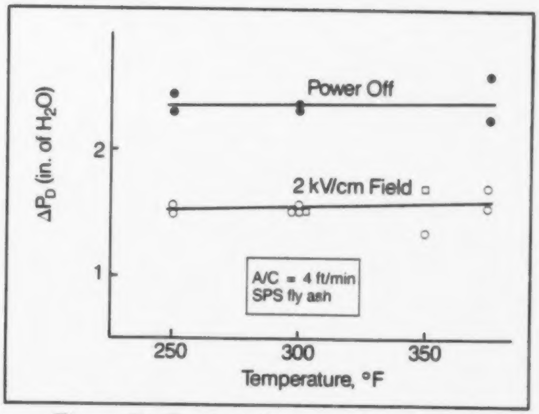


Figure 7. Extended temperature range data (reverse air, SPS fly ash).

ture operation of the Harrington Station pilot unit should not compromise the conclusions.

ESFF APPLIED TO SPRAY DRYER SOLID BYPRODUCT

With increasing usage of spray drying techniques to contact boiler emissions with an alkaline solution or slurry to remove sulfur oxides, the byproduct of this process, most often removed from the flue gas by a baghouse, is an attractive candidate dust for ESFF. The spray dryer byproduct normally differs from fly ash in electrical resistivity. It is collected at a low temperature, on the order of 150° F (66° C), where the reaction rate is favored, and it contains a larger amount of water vapor than does common fly ash.

Spray dryer solid byproduct was supplied by Joy Manufacturing Company. The material was collected from the baghouse at their Riverside Plant where the operating temperature was 160° F (71° C). They also supplied typical electrical resistivity information which indicated that the material has a resistivity of about 10^8 ohm-cm with 12 percent moisture. By extrapolation to this low temperature, EPA measurements of resistivity also yielded about the same result. The EPA laboratory measurements are shown in Figure 8. The byproduct resistivity at two moisture levels and the SPS fly ash resistivity at one level are shown. A reference plot of SPS dry fly ash is superimposed, and several points from the Joy resistivity determinations are also shown. The important observations are the big changes in resistivity of the byproduct with moisture and temperature. This has a bearing on the ESFF measurements to be described below.

COMPARISON WITH FLY ASH

The reason for applying ESFF to the spray dryer byproduct was simply to determine if filtration of this material is as amenable to electrical enhancement as is fly ash. Initially the byproduct was injected into the air stream at the same temperature, A/C, and grain loading typically used for fly ash. The results from this first attempt were not noteworthy: little enhancement was observed. Further experiments, in which the temperature of the filtration was lowered and moisture injected along with the dust, produced an ESFF effect that equalled and surpassed the fly ash results. Table 1 gives the results of the comparison of byproduct and fly ash. The ESFF effect, given as PDR (pressure drop ratio), is the slope of the ΔP curve for the electrically enhanced filtration divided by the slope of the ΔP curve for filtration under the same conditions but with the electrical power turned off. The data suggest that a combination of low temperature and moisture has a synergistic beneficial effect which has not been seen before. By just lowering the temperature to 200° F (93° C) (which allows an increase in the maximum electrical field prior to corona onset), the ESFF effect in the spray dryer byproduct approaches that of the fly ash at 300° F (149° C). Further reduction in the resistivity can be accomplished by moisture addition, and still higher fields can then be maintained, and consequently a lower PDR is achieved. These observations showed spray dryer byproduct to be compatible with ESFF. Other experiments were then carried out to determine the magnitude of electrical enhancement

TABLE 1. ESFF RESULTS: COMPARISON OF FLY ASH AND SPRAY DRYER BYPRODUCT

Baghouse	Dust Type	Water Added, %	Temperature, °F	Electric Field, kV/cm	Loading, grains/ft^3	Effect (PDR)
Pulse Jet	Fly ash	0	300	6	2.5	.4-.5
	Byproduct	0	300	6	2.5	1
	Byproduct	0	200	6.5	2.5	.67
	Byproduct	6	300	7	2.5	.4
	Byproduct	5	200	7.5	2.5	.26
Reverse Air	Fly ash	0	300	2	2.5	.5-.6
	Byproduct	0	300	2	2.8	.85
	Byproduct	0	200	2	2.8	.74
	Byproduct	5	250	2	2.3	.6
	Byproduct	5	200	2.5	2.3	.29

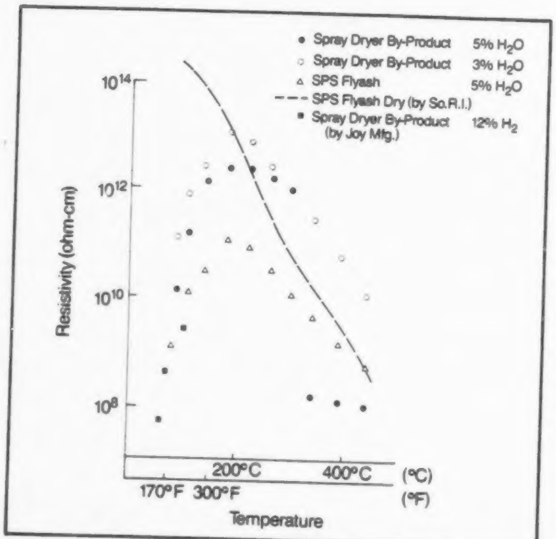


Figure 8. Fly ash electrical resistivity.

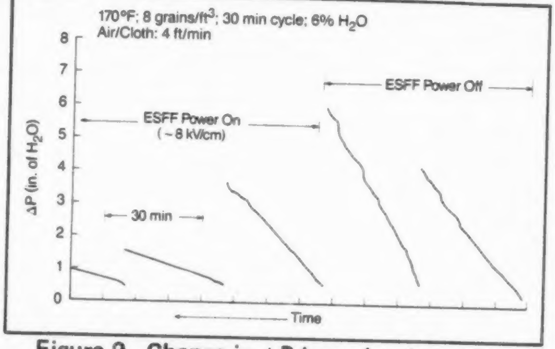


Figure 9. Change in ΔP brought about by turning the electric field on (pulse jet, spray dryer byproduct).

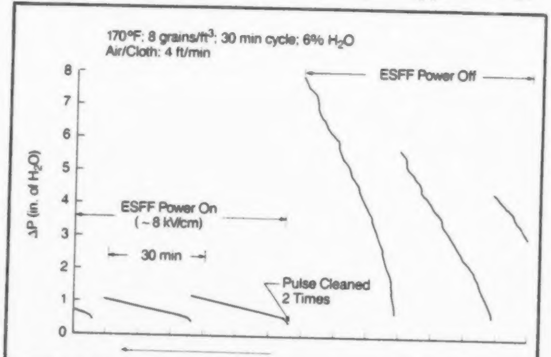


Figure 10. Change in ΔP brought about by turning the electric field on (pulse jet, spray dryer byproduct, with two cleaning pulses).

under conditions approximating realistic operating conditions and are described next.

EXPERIENCE WITH SPRAY DRYER SOLID BYPRODUCT

EPA in-house application of ESFF to a spray dryer solid byproduct supplied by Joy Manufacturing Company has shown that this material can benefit from electrical enhancement. The correlations of ESFF performance with temperature and electrical field and its dependence on dust electrical resistivity have promoted basic understanding of the ESFF mechanism.

It has already been mentioned that ESFF was not very effective when applied to spray dryer byproduct at fly ash operating temperatures and common fly ash grain loadings. One reason appears to be the very high electrical resistivity of dry byproduct at 300° F (149° C). Increasing the moisture level to near actual operating conditions decreases the byproduct resistivity significantly. A combination of high moisture and low temperature in the range of 150° to 170° F (66° to 77° C) resulted in PDRs in the 0.15 to 0.2 range--well below the range typically observed even for fly ash. Furthermore, this response to ESFF held up at the high grain loadings associated with spray dryer outlet conditions.

The extremely good response of spray dryer byproduct to ESFF is perhaps best illustrated by pressure drop comparisons taken directly from recorder charts. Figures 9 and 10 are reproduced from the pulse-jet recorder. Figure 9 shows two ΔP curves with power off followed by three curves with power on (ESFF). The first curve after a field has been applied to the electrodes is still under the influence of the conventional cycle which has just preceded turning power on. One pulse of compressed air has not been sufficient to completely clean the bag which, without the electric field, is uniformly covered with dust. Two pulses did clean sufficiently as shown by the almost complete recovery of the first ESFF curve in Figure 10.

A photograph of dust deposited under ESFF conditions, Figure 11, helps to explain why dust deposition with the electric field on gives a lower pressure drop buildup with time. As seen in Figure 11, dust deposition is along the electrodes. This accounts for the lower ΔP because a nearly clean fabric path is provided which offers little resistance to gas flow. The highly nonuniform dust distribution results in a much lower pressure drop than a uniform distribution of the same quantity of dust. The nonuniform dust layer also proves easy to remove during cleaning, reducing the residual ΔP creep characteristic of most baghouse operations.

The ΔP curves for conventional filtration (power off) in Figures 9 and 10 are not reproducible or consistent. Experience with the pulse-jet unit indicated that, under this high grain loading, only a limited number of filtration cycles could be performed before the pressure drop curve would exceed the chart limit. This is a consequence of inadequate cleaning (5); increasing the cleaning energy might have kept the ΔP on the chart. Under ESFF, however, these same cleaning pulses provided sufficient cleaning energy to give a stable reproducible ΔP plot. Because PDR is based on a comparison of ESFF and conventional ΔP's, the PDR's reported herein are based only on the average of a limited number of conventional cycles.

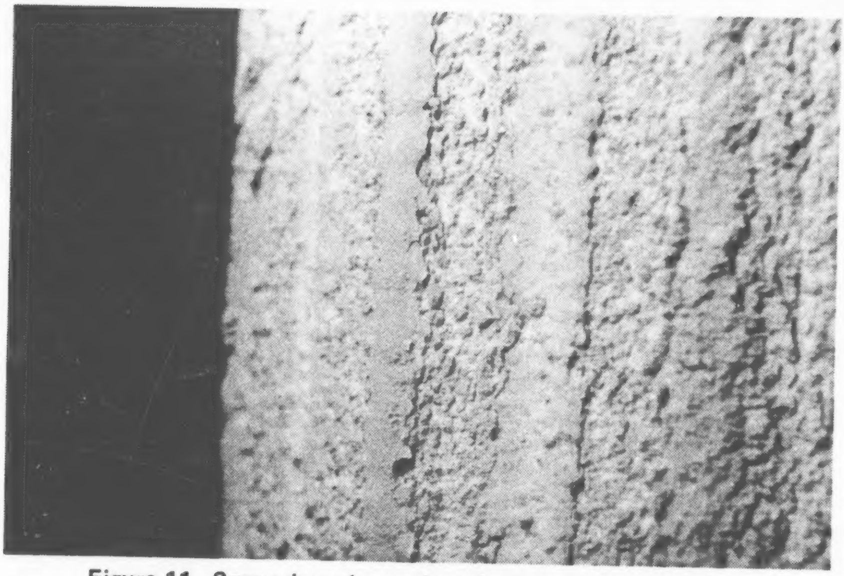


Figure 11. Spray dryer byproduct deposit on electrified bag (pulse jet cleaning).

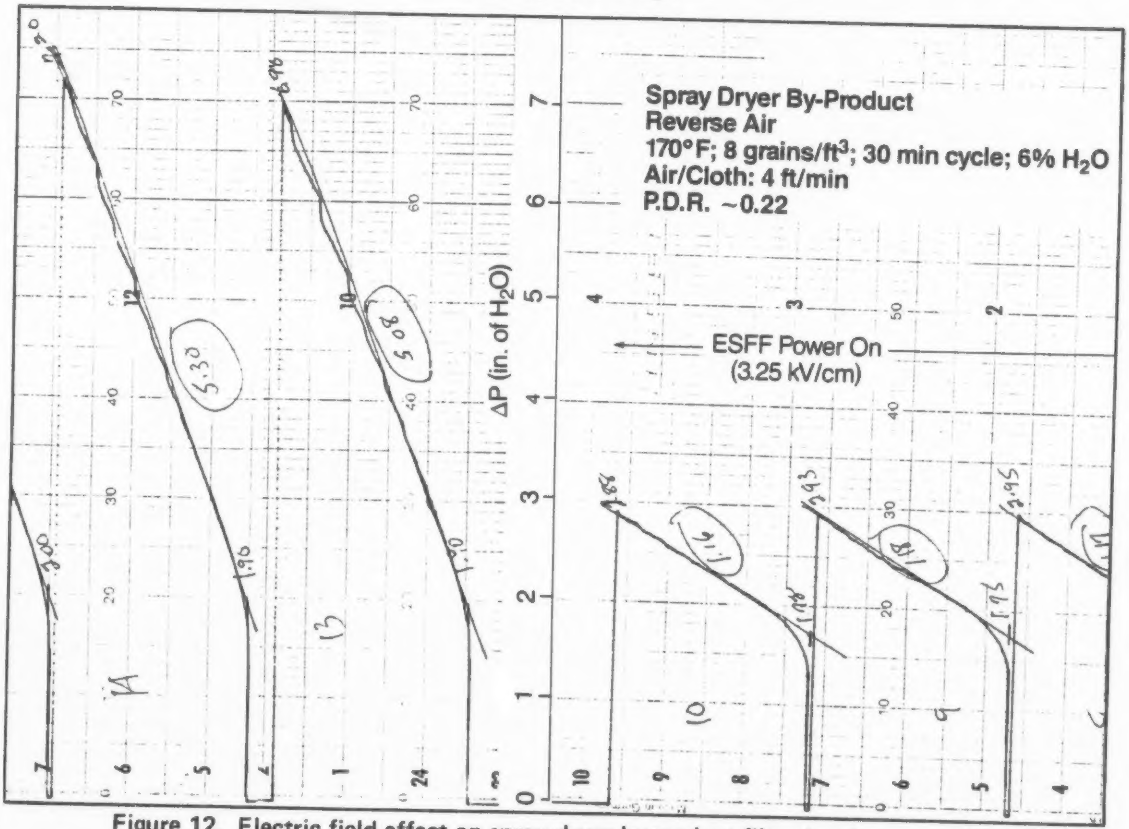


Figure 12. Electric field effect on spray dryer byproduct filtration (reverse air).

Figure 12 is a reverse-air recorder chart comparison similar to the one for pulse-jet just reviewed. The PDR values are extremely good here also, being in the range of 0.2 to 0.3. Dust deposition along the electrodes is not readily observed because of the inside-the-bag nature of filtration. However, there is no reason to think that the mechnism for achieving a low pressure drop differs from that of the pulse-jet case. The record of ΔP indicates that, for the cycling shown in Figure 12, there is sufficient cleaning energy to remove the uniform layer of dust under conventional filtration because the ΔP curve is reproducible. The pressure drop rose above the chart limit for the power-off condition only when the filtration cycle time was doubled.

The striking difference between the power-on and power-off curves of Figures 9 and 10 (pulse-jet) and Figure 12 (reverse-air) can be appreciated by comparing them with similar curves for fly ash (Figure 13, reverse-air). The ESFF effect, while clearly evident, is much less pronounced in the Figure 13 data which yield a PDR typically in the 0.4 to 0.6 range.

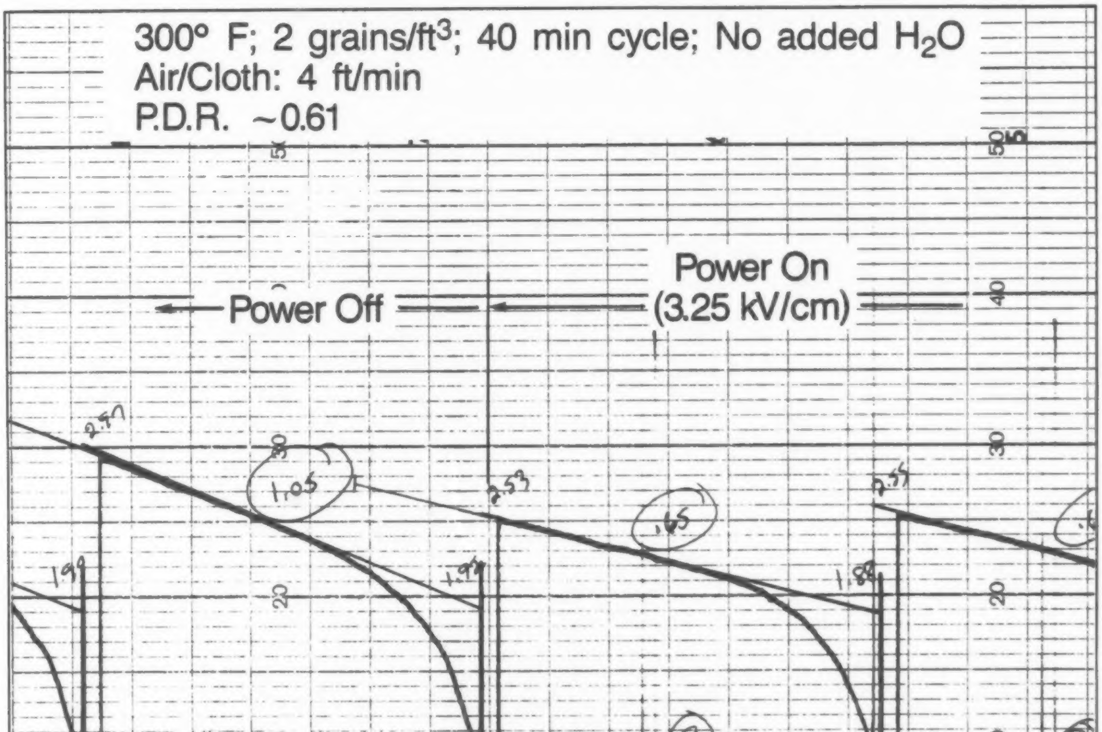


Figure 13. Electric field effect on SPS fly ash filtration.

VERIFICATION OF ESFF MECHANISM

The nonuniform dust deposition pattern, alluded to in describing the lower ΔP_D, and the ease of cleaning were readily observed through the port on the pulse-jet baghouse. The nonuniform dust deposition pattern, which was emphasized by the large grain loading in the case of spray dryer byproduct, is most likely a consequence of relatively long range coulomb forces exerted on the dust (6). Charged particles are attracted to either the positive or negative electrode with more depositing on the electrode of polarity opposite to the charge on the majority of particles. Other mechanisms (7, 8) may play a role, but the nonuniform deposition clearly dominates. Figures 11 and 14 show the buildup of dust along the electrodes. That the dust can be adequately removed from the bag, as evidenced by a stable, low value of residual ΔP, also contributes to the effectiveness of ESFF.

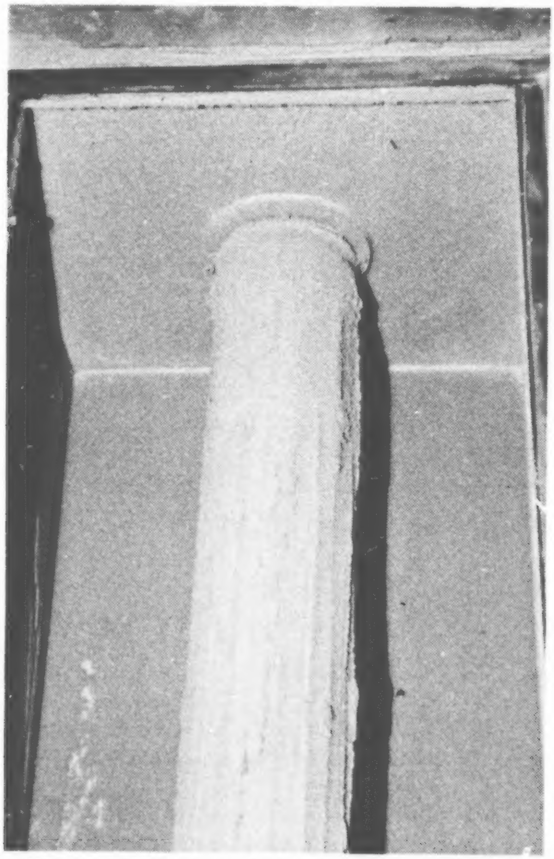


Figure 14. Spray dryer byproduct deposition pattern (pulse jet).

MAXIMIZING THE ESFF EFFECT

It has been concluded that the spray dryer byproduct, because of the conditions under which it is filtered, is a likely candidate for ESFF. Although the dust has a very high electrical resistivity when dry, a combination of low operating temperature and high moisture lowers this resistivity into a favorable range for ESFF. Both high moisture and low temperature are necessary. Figure 15 shows a pulse-jet response to a reduction in filtration temperature versus a reduction in temperature with added moisture. The lower PDR response to temperature is measurable, but it takes the added moisture to observe the impressively low values of PDR. There is no intended implication that electric field is held constant in Figure 15. The Figure 15 data were collected at "maximum" allowable electric field--that value of electric field just below corona onset. This value of field increases with increasing moisture and decreasing temperature. Thus, the low-temperature, high-moisture data of Figure 15 were collected under higher values of applied electric field than the low-moisture, high-temperature data.

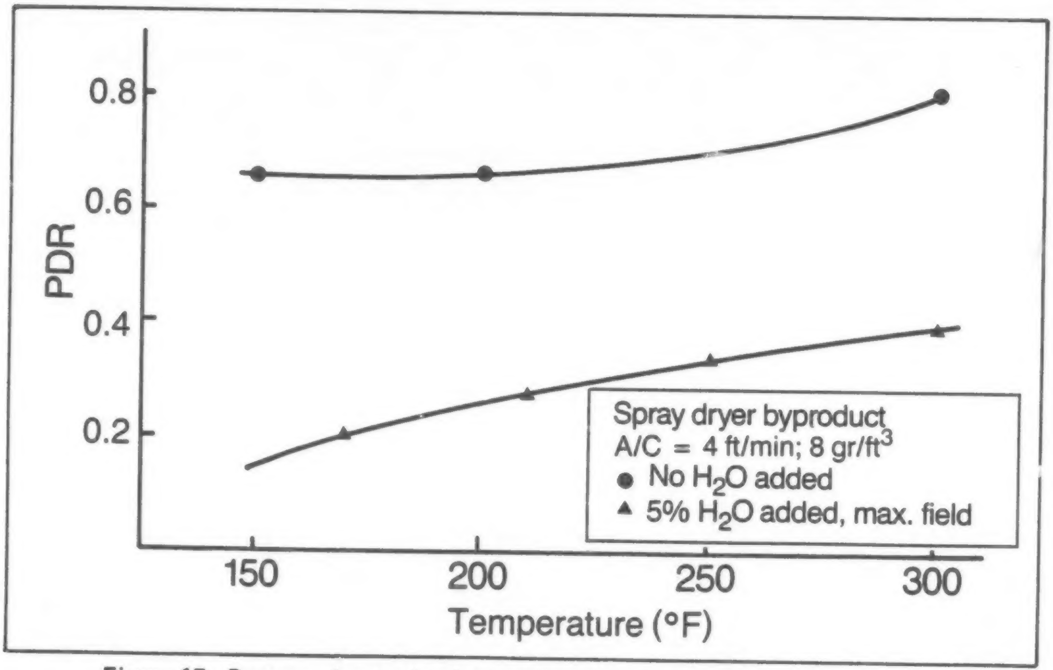


Figure 15. Pressure drop reduction attributable to reduced temperature and increased moisture (spray dryer byproduct, pulse jet).

This field dependence is demonstrated for pulse-jet operation in Figure 16 and for reverse-air operation in Figure 17. It is interesting to note that an applied electrode voltage of approximately 11 kV (5.5 kV/cm) and approximately 4 kV (2 kV/cm field) for pulse-jet and reverse-air, respectively (at this temperature and moisture condition), gives an ESFF result comparable to SPS fly ash at 300° F (149° C) and low moisture. Data are not available to confirm this, but it is surmised that electrical resistivities are nearly the same for the two products at these conditions.

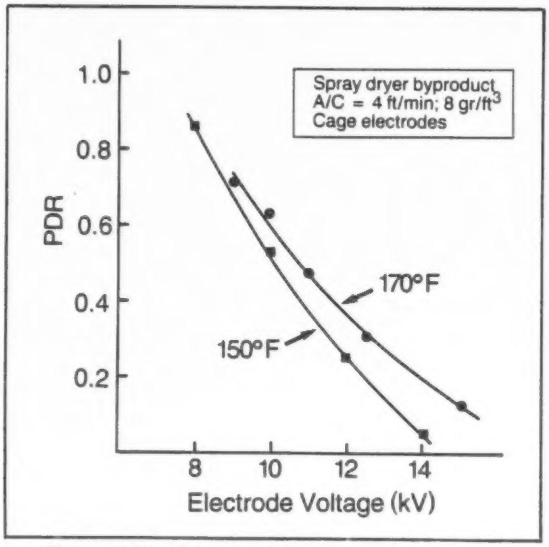


Figure 16. Field dependence of pressure drop reduction (spray dryer byproduct, pulse jet).

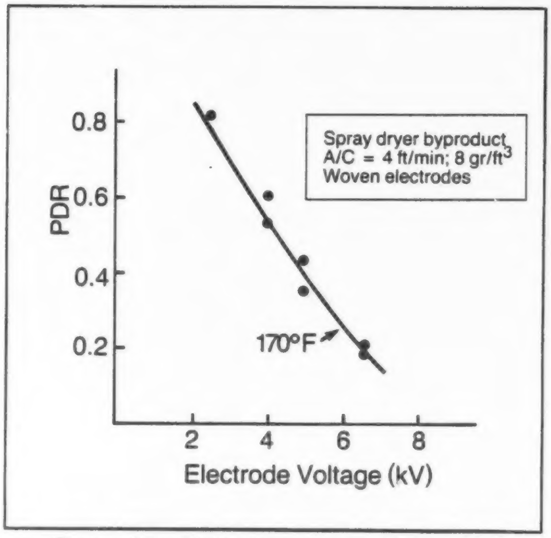


Figure 17. Field dependence of pressure drop reduction (spray dryer byproduct, reverse air).

CONCLUSIONS

New, versatile high-temperature baghouses recently installed in EPA's Air and Energy Engineering Research Laboratory at Research Triangle Park have already proven their worth in evaluating ESFF at varying temperatures and in testing ESFF on spray dryer byproduct. The use of this equipment to investigate other aspects of fabric filtration will be reported later. Conventional and nonconventional fabrics have been tested, and alternative electrode configurations have been studied in a continued search for lower cost and more easily maintained fabric filtration.

METRIC CONVERSIONS

1 in. of H_2O = 249 Pa

1 grain/ft^3 = 2.29 g/m^3

1 ft/min = 0.5 cm/s

REFERENCES

1. Lamb, G. E. R., Costanza, P. A., and O'Meara, D. J., Electrical Stimulation of Fabric Filtration. Part II: Mechanism of Particle Capture and Trials with a Laboratory Baghouse. Textile Research Journal 48. No. 10, Oct. 1978, pp. 566-573.

2. Masuda, S., Nakatani, H., and Mizuno, A. Boxer-Charger Mark III and its Applications in ESP's. In: Third Symposium on the Transfer and Utilization of Particulate Control Technology. Vol. II, EPA-600/9-82-005b (NTIS PB83-149591), July 1982, pp. 380-389.

3. Greiner, G. P., Furlong, D. A., VanOsdell, D. W., and Hovis, L. S. Electrostatic Stimulation of Fabric Filtration. JAPCA 31, No. 10, Oct. 1981, pp. 1125-1130.

4. Chambers, R. L., Spivey, J. J., and Harmon, D. L. ESFF Pilot Plant Operation at Harrington Station. Paper 86, Fifth Symposium on the Transfer and Utilization of Particulate Control Technology. Kansas City, MO, Aug. 1984.

5. Leith, D., and Ellenbecker, M. J. Theory for Pressure Drop in a Pulse Jet Cleaned Fabric Filter. Atmospheric Environment. No. 14, 1980, pp. 845-852.

6. Chiang, T. K., Samuel, E. A., and Wolpert, K. E. Theoretical Aspects of Pressure Drop Reduction in a Fabric Filter with Charged Particles. In: Third Symposium on the Transfer and Utilization of Particulate Control Technology. Vol. III, EPA-600/9-82-005c (NTIS PB83-149609), July 1982, pp. 250-260.

7. Morris, B. A., Lamb, G. E. R, and Saville, D. A. Modeling Studies of Pressure Drop Reduction in Electrically Stimulated Fabric Filtration. Paper 69. Fifth Symposium on the Transfer and Utilization of Particulate Control Technology. Kansas City, MO, Aug. 1984.

8. VanOsdell, D. W., Donovan, R. P., and Hovis, L. S. Flow Resistance Reduction Mechanisms for Electrostatically Augmented Filtration. Paper 70, Fifth Symposium on the Transfer and Utilization of Particulate Control Technology. Kansas City, MO, Aug. 1984.

COMPARISON OF PILOT PLANT DATA WITH PREDICTIONS FROM THE U.S. EPA SPRAYMOD COMPUTER PROGRAM

by

Wayne T. Davis
Gregory D. Reed
George P. Partridge
Civil Engineering Department
The University of Tennessee
Knoxville, TN 37996

and

Theodore G. Brna
U.S. EPA, AEERL
Research Triangle Park, NC 27711

ABSTRACT

The use of computer models allows the user to predict conditions which may not have been tested in spray dryer/fabric filter applications. This paper presents a preliminary comparison of the U.S. EPA developed SPRAYMOD computer program with a series of five different data bases collected at the University of Tennessee on a 1000 cfm spray dryer utilizing a $Ca(OH)_2$ slurry. Results are presented from a variety of operating conditions for the spray dryer without recycle, including:

SO_2 concentration:	1000 - 3100 ppm
Inlet temperature:	138° - 177°C (280-350°F)
Approach to saturation (AT):	9° - 20°C (16-35°F)

Conclusions regarding the ability of the program to predict the actual behavior of the pilot plant are: (1) with the proper choice of a reaction rate coefficient, the plug flow model reliably predicted the performance for low SO_2 concentrations (1000-1100 ppm) and approaches to saturation in the range of 16-35°F, and (2) the model tended to overpredict performance for higher SO_2 concentrations (2000-3000 ppm).

This paper has been reviewed in accordance with the U.S. Environmental Protection Agency's peer and administrative review policies and approved for presentation and publication.

COMPARISON OF PILOT PLANT DATA WITH PREDICTIONS FROM THE U.S. EPA SPRAYMOD COMPUTER PROGRAM

INTRODUCTION

This paper presents a preliminary comparison of the U.S. EPA developed SPRAYMOD computer program with a series of five different data bases collected at the University of Tennessee on a 1000 cfm spray dryer utilizing a $Ca(OH)_2$ slurry. Curves of measured SO_2 removal efficiency across the spray dryer versus stoichiometric ratio are compared to predicted curves using SPRAYMOD. The preliminary comparisons identify the parameters in the model which are likely to be the most sensitive.

EXPERIMENTAL DESCRIPTION OF PILOT PLANT

Figure 1 is a schematic diagram of the pilot test facility which was constructed in 1980 on a slipstream of the University of Tennessee's Riley stoker-fired boilers. The original system consisted of a slaking system control room, a 7 ft diameter spray drying chamber with a variable speed Stork-Bowen AA-6 Spray Machine (5000-21,000 rpm), and a 6 in. diameter centrifugal atomizer with six nozzle inserts. A low energy shaker baghouse with four 12 in. x 32 ft fiberglass filter bags (bottom entry) was utilized downstream from the spray dryer for particulate control. In July 1984, the shaker baghouse was replaced by a 36 bag pulse jet collector (off-line cleaning) with 4.5 in. x 12 ft fiberglass filter bags.

A continuous pneumatic recycle system was installed in 1983 for use in recycle testing. This system provided a continuous partial recycle of the dried product collected in the spray dryer. Since the baghouse was periodically cleaned (rather than continuously cleaned), the baghouse product was pneumatically conveyed to a cyclone separator, collected in a screw feeder, and partially recycled back into the slurry feed as needed. The lime feed, spray dryer recycle, and baghouse recycle were mixed in an intermediate mixing tank, and the resulting slurry screened through a 420 μm vibrating screen prior to delivery to the atomizer. This technique was found to be necessary for successful operation at solids concentrations of 15-35% solids to prevent clogging of the slurry lines and atomizer nozzles due to the small volumetric flow rate of the delivery system (less than 0.30 gpm with 0.25 in. ID lines). Further it was found that it was necessary to continuously recycle the lime slurry between the final slaking room mixing tank and the recycle tanks located at the spray dryer in order to prevent plugging in the transport lines when the atomizer feed rate was terminated. In the testing reported here, no recycle data are included; hence, the recycle transport system was not operated.

System instrumentation provided for continuous monitoring of the static pressure, temperature, and SO_2 concentrations at the inlet and outlet of both the spray dryer and the baghouse. Sampling locations for these are indicated in Figure 1 by the symbols P, T, and S, respectively. Continuously extracted samples of flue gas were transported through heat-traced lines to one of three analyzers: two Lear-Siegler SM810 analyzers and a TECO Model 800/Model 40 gas conditioner/analyzer. Depending on the availability of analyzers, data were acquired simultaneously at multiple ports or sequentially. Wet bulb temperatures were measured at all three ports by a hand-held wetted-wick thermometer and/or thermocouple (symbol TW in Figure 1).

Temperature was controlled by a dilution air damper located upstream of the inlet sampling ports of the spray dryer. The concentration of SO_2 was maintained by controlled supplemental injection of SO_2.

TEST PROCEDURES FOR SPRAY DRYER TESTING

All data reported here represent individual tests of 45 minutes to 1 hour duration. A single 1 hour evaluation consisted of monitoring 18 different parameters at the pilot plant including atomizer disk speed, SO_2 concentrations at Ports 1-3, temperature at Ports 1-3, available CaO (by HCl titration), orifice pressure drop for flow rate, static pressures at Ports 1-3, lime slurry flow rate, barometric pressure, wet bulb temperatures at Ports 1-3, and ambient temperature. Air in-leakage rate and flue gas molecular weight (Orsat analyzer) were conducted periodically. In addition to these measurements, analyses were conducted in the laboratory to determine the size distribution of the slaked $Ca(OH)_2$ particles in the slurry, and the moisture content of the reacted products collected in the spray dryer and fabric filter collector hoppers.

BRIEF DESCRIPTION OF THE SPRAYMOD COMPUTER MODEL

The SPRAYMOD computer program is a simulation program which mathematically models the removal of SO_2 in a spray dryer. The model developed by Research Triangle Institute (1) under contract to EPA is a simultaneous heat and mass transfer model which predicts the simultaneous evaporation of a sorbent droplet and absorption/reaction of SO_2 in the evaporating droplet. Further the model includes the dissolution kinetics of the $Ca(OH)_2$ particles within the droplet. Since the detailed description of SPRAYMOD has been reported elsewhere (1) it will not be discussed here, except as needed for clarification of the input parameters to the model.

INPUT TO SPRAYMOD

Table 1 summarizes the input parameters required to execute the computer simulation. These parameters include inlet gas conditions, operating variables, sorbent properties, and program variables.

Inlet Gas Conditions

The required gas conditions which are input to the model include the spray dryer inlet gas temperature (°C), the inlet water content of the flue gas (% by volume), the inlet SO_2 concentration (in ppm on a wet basis), and the flue gas molecular weight (on a dry basis).

Operating Conditions

The operating parameters include the approach to the adiabatic saturation temperature (°C) at the spray dryer outlet, the stoichiometric ratio of $Ca(OH)_2$ (moles of $Ca(OH)_2$/mole of inlet SO_2), mass fraction of reactive sorbent in fresh solids (1.0 if no recycle), recycle ratio (zero for no recycle), mass fraction of active sorbent in recycle (zero for no recycle), residence time (seconds) of the gas phase based on the volume flow rate at the outlet of the spray dryer, the type of mixing condition in the spray dryer (complete mix versus plug flow), and an assumed spray dryer efficiency. This latter parameter is only utilized in the complete mix option as a starting point for the calculations.

Sorbent Properties

Sorbent properties required as input to the model include the inlet droplet diameter (µm), the inlet droplet temperature (°C), form of sorbent (slurry versus solution), sorbent mean particle diameter (µm), molecular weight of sorbent (74 g/mole for $Ca(OH)_2$), the density of solid sorbent (2.2 g/cm^3), critical moisture content (XC), equilibrium moisture content (XE), and the dry sorbent reaction rate coefficient (KR).

The inlet droplet diameter used in this study for the pilot plant data was a value calculated utilizing the equation developed by Masters (2):

$$DVS = \frac{1.4 \times 10^4 (M)^{0.24}}{(Nd)^{0.83} (nh)^{0.12}} \qquad (1)$$

where:

DVS = Sauter mean diameter (µm)
M = mass feed rate (kg/h)
N = atomizer wheel speed (rpm)
d = atomizer wheel diameter (m)
n = number of vanes
h = atomizer wheel vane height (m)

For the pilot plant data described here, M = 37-57 kg/h, N = 17000 rpm, d = 0.15 m, n = 6, and h = 0.0066 m. The resulting range of droplet diameters is 73-81 µm.

The solvent particle diameter, based on measurement of the particle size distribution in the laboratory, was found to be 2.9-3.4 µm (volume mean diameter) with a geometric standard deviation of 1.5-1.8. An average value of 3.25 µm was used here. This compares with data reported by Stevens et al. (3) in which the range was reported to be 3.5-4.4 µm.

The equilibrium moisture content of the drying product is defined as the water content that would exist if the flue gas were at 100% saturation (100% relative humidity). This value was approximated by assuming that the moisture content of the product was linearly related to the relative humidity. Figure 2 shows a graph of water content of product versus stoichiometric ratio for pilot plant data collected in the UT system (Data Set 4). For the condition tested (8.9-10°C approach to saturation), the moisture content was approximately 1-11%, with an average value of approximately 3%. Previous test data on a full-scale spray dryer have shown moisture contents of 2-3% at the same approach to saturation (4). The relative humidity for these conditions was 40-50%; using 50% relative humidity and 3% moisture content, the equilibrium moisture content was estimated to be approximately 6% for the data evaluated in this study.

The critical moisture content of the product, defined as the condition in SPRAYMOD at which the model switches to a "dry particle stage" reaction, was chosen as 50% (1). Further, the dry sorbent reaction rate coefficient (KR) was initially assigned a value of zero. This value, as discussed by Damle and Sparks (1), is a function of the particle moisture content and will be discussed further in this paper in light of the results from data analyses.

Tables 1 and 2 summarize the data and inputs utilized in the comparisons conducted in this paper. While the final analyses of these data will involve comparisons of predicted versus measured efficiency for each data point in all of the data sets, a preliminary analysis was conducted by comparing the measured efficiency versus stoichiometric ratio curve for each data set to the approximate predicted efficiency versus stoichiometric ratio curve for the data sets. In this manner the number of preliminary analyses was minimized by only requiring a selected few calculations for each curve.

PRELIMINARY SENSITIVITY ANALYSES

Prior to conducting the above analyses, a partial sensitivity analysis was conducted of SPRAYMOD to aid in determining whether the backmix or plug flow model would be utilized in the data comparisons and to determine which operating parameters would result in significant changes in the predicted efficiency. A baseline condition was chosen about which each input parameter was varied. The baseline conditions are listed in Table 2. Figure 3 shows a comparison of the predicted efficiency for the model assuming the backmix and plug flow reaction schemes. The values indicate that the backmix and plug flow options show the efficiency to be relatively insensitive to most operating parameters. Further, these predicted efficiency values (for the spray dryer with lime only) were observed to be consistently higher than the values observed in the pilot plant tests to be analyzed.

The plug flow model showed significant dependence on the reaction rate coefficient. The predicted efficiency changed significantly as the reaction rate coefficient was varied from 0 to 10^7cm^3/mole–sec as shown in Figure 3. It was also noted that, for the condition in which the reaction rate coefficient was 10^7cm^3/mole–sec, the plug flow model was relatively insensitive to other operating and gas characteristics such as residence time, approach to saturation, droplet diameter, and lime particle size since the utilization was relatively high (greater than 85%). However, when KR was set equal to zero, resulting in a significant reduction in efficiency (plug flow model only), the model became much more sensitive to the other variables.

Based on the above observations from the sensitivity analyses and the reported and observed sensitivity of the measured efficiencies to lime particle size and approach to saturation, the preliminary analyses reported here were conducted with the plug flow model utilizing a KR value of zero.

COMPARISON OF SPRAYMOD TO SPECIFIC DATA SETS

Analyses were conducted using the operating and gas parameters shown in Table 2. In total, five different data sets were utilized for the preliminary analyses. These included efficiency versus stoichiometric ratio curves at the following approximate concentrations of SO$_2$/inlet temperatures/approach to saturations:

Data Set 4 (Figure 4): 1100 ppm/150°C (302°F)/8.9–10°C (16–18°F)
Data Set 5 (Figure 4): 1100 ppm/148°C (298°F)/11°C (20°F)
Data Set 1 (Figure 5): 1000 ppm/172°C (342°F)/19.4°C (35°F)
Data Set 2 (Figure 6): 2000 ppm/171°C (340°F)/19.4°C (35°F)
Data Set 3 (Figure 6): 3000 ppm/169°C (336°F)/19.4°C (35°F)

DISCUSSION

As shown by the comparisons, the model tended to predict the lower concentration data (1000-1100 ppm) at the various approaches to saturation. However, it significantly overpredicted the higher concentration data (2000-3000 ppm). Another observation which was generally observed was that the predicted curves were consistently less sensitive to stoichiometric ratio (SR) than the observed data for greater than 1.0 SR.

It was hypothesized that the relatively flat efficiency at SR's greater than 1.0 would be affected by the value of KR, the reaction rate coefficient. This was tested by conducting a brief sensitivity analysis of the effect of KR on efficiency. Figure 7 shows a graph of efficiency versus stoichiometric ratio for KR ranging from 0 to 10^7 for the baseline conditions in Table 2. Note that the reaction rate coefficient does not affect the efficiency until the KR exceeds a value of 10^5 cm^3/mole-sec. Further, the effect of the choice of the KR value is more pronounced at the higher SR values than at SR values less than 0.9. This leads to the realization that the ability of the model to accurately predict the observed efficiency depends to a large extent on the user's choice of a reaction rate coefficient. As is emphasized by Damle and Sparks (1) the reaction rate coefficient depends upon the diffusivity of SO_2 in the solid material, the particle size, and the moisture content of the droplet/particle. Since the lime particle size can be measured, the primary unknown factor affecting KR is the moisture content. At the present time, the value of KR has not been measured and included in the model. Davis and Keener (5) have measured the reaction rate coefficient, KR, of dry $Ca(OH)_2$ in a bench-scale reactor and obtained a value of approximately 25 cm^3/mole-sec at a temperature of 140°F. However, it is expected that KR would be orders of magnitude higher at 1-10% moisture content. Based on the evaluation included here, it is recommended that a value of KR = 0 be assumed until further information becomes available.

Additional sensitivity analyses were conducted on Data Sets 3-5 to determine the sensitivity of the lime particle size on the model-predicted values. It was observed that if a lime particle size of 4.2 m had been used (as opposed to the 3.25 µm actually measured) the predicted efficiencies of the 1000, 2000, and 3000 ppm data would have been 58, 63, and 63.4%, respectively, at a stoichiometric ratio of 1.5. This would have resulted in a significant underprediction of the 1000 ppm data and a closer simulation of the 2000-3000 ppm data. However, it was observed that the lime slurry particle size data collected in this study were all similar since all slaking was done in the same manner, eliminating this as a possible explanation for the inability of the model to predict the performance for both high and low SO_2 concentrations. The sensitivity of the model to lime particle size does suggest, however, that further study is needed to determine if the dissolution equation is the source of the observed inability to predict both low and high SO_2 concentration data. The effect of increased slurry concentrations at the higher SO_2 concentrations may be affecting the diffusivity of the calcium ions, resulting in decreased dissolution rate. Further study is currently in progress.

CONCLUSIONS

Preliminary analyses have shown that SPRAYMOD has the capability of predicting SO_2 removal efficiency of a pilot lime spray dryer for low SO_2 concentrations (1000-1100 ppm) for approaches to saturation in the range of 9-20°C (16-35°F) utilizing the plug flow model. However, the ability to predict measured efficiencies is highly dependent on the user's choice of a reaction rate coefficient, KR. Further development work is needed to better define the value of KR as a function of lime particle size and/or product moisture content. Since the model conducts a stepwise integration as the product moisture is evaporated, the inclusion of a KR which is a function of moisture content could be easily adapted into the model if data were available. In the absence of this modification, it is recommended that a KR value of zero be used.

Based on the preliminary comparisons of the model to measured data for higher SO_2 concentrations (2000-3000 ppm), the model tended to overpredict the measured values. Further study is in progress to determine if this overprediction can be corrected by modifying the dissolution equations in the model to account for potential effects of higher slurry concentrations on the dissolution of the lime particles.

REFERENCES

1. Damle, A. S. and L.E. Sparks. Modeling of SO_2 Removal in Spray Dryer Flue Gas Desulfurization System. Presented at the Fifth Symposium on the Transfer and Utilization of Particulate Control Technology, Kansas City, MO., August 27-30, 1984.

2. Masters, K. Spray Drying Handbook. John Wiley & Sons, 1979, pp. 179-180.

3. Stevens, N.J., G.B. Manavizadeh, G.W. Taylor, and M.J. Widico. Pilot-scale Parametric Testing of Spray Dryer SO_2 Scrubber for Low-to-Moderate Sulfur Coal Utility Applications. EPA-600/7-84-045 (NTIS No. PB84-175959), March 1984, pp. 93-104.

4. Kelly, M.E. and M.A. Palazzolo. Status of Dry SO_2 Control Systems: Fall 1982. EPA-600/7-83-041 (NTIS No. PB83-247585), August 1983, pp. 105-106.

5. Davis, W.T. and T.C. Keener. Chemical Kinetic Studies on Dry Sorbents. DOE/FC/10184-1 (NTIS No. DE82012765), 1982, p. 38.

ACKNOWLEDGEMENTS

The authors would like to acknowledge the assistance of Ashok Damle, Research Triangle Institute, who made many very helpful suggestions in the application/interpretation of the SPRAYMOD program.

Also the authors acknowledge the previous research support provided by Flakt, Inc., Knoxville, TN, from which the majority of the test data were obtained, and U.S. EPA in supporting the program comparisons reported here.

TABLE 1. INPUT TO SPRAYMOD

Input Data Set - PROGRAM SIMULATION

Inlet Gas Conditions -

Inlet Gas Temperature (TI) - 160°C
% Water in Inlet Gas (WI) - 6.5%
Inlet S02 Concentration (SI) - 800 ppm
Molecular Weight of Dry Inlet Gas (MG) - 28.9

Operating Parameters -

Approach to Saturation (AT) - 15°C
Stoichiometric Molar Ratio (SR) - 1
Mass Fraction of Sorbent in Fresh Solids (MF) - 1
Recycle Solids Ratio by Mass (RR) - 0
Mass Fraction of Sorbent in Recycle (MR) - 0
Residence Time of Gas Phase (RT) - 10 s
Flow System in Spray Dryer - OP = 1 (Plug flow) - OP = 2 (Backmix)
Assumed Spray Dryer Efficiency (P1) - 50%

Sorbent Properties -

Inlet Droplet Diameter (DO) - 50µm
Inlet Droplet Temperature (TW) - 60°C
Form of Sorbent - Slurry (OS = 1)
Sorbent Particle Diameter (DP) - 4µm
Molecular Weight of Sorbent (MS) - 74
Density of Solid Sorbent (SG) - 2.2 g/cm^3
Critical Moisture Content (XC) - 50%
Equilibrium Moisture Content (XE) - 3%
Dry Sorbent Reaction Rate Coefficient (KR) - 1E+07 cm^3/gmole-s

Program Parameters -

Maximum Allowable Time Step (MT) - 0.1 s
Time Between Printouts (PT) - 0.5 s
Time Step Control Parameter (SC) - 0.01 s

TABLE 2. SUMMARY OF SENSITIVITY ANALYSES FOR THE BASELINE AND PILOT PLANT DATA SETS[a]

Parameter	Baseline	DATA SET*				
		1[b]	2[b]	3[b]	4[c]	5[c]
TI (°C)	149	172	171	169	150	148
WI (%)	5	5.6	6.9	6.6	4.3	5.5
SI (ppm)	1100	1012	1986	2893	1103	1075
MG	28.6	29.4	29.4	29.4	29.4	29.4
AT (°C)	11.1	19.4	19.4	19.4	8.9–10	11.1
SR	1.0	0.6 to 2.0				
MF	1.0	1.0	1.0	1.0	1.0	1.0
RR	0	0	0	0	0	0
MR	0	0	0	0	0	0
RT (sec)	17	17.5	16.4	17.6	16.6	19.5
OP	0,1	plug flow (OP = 1)				
DO (μm)	80	77	78	79	80	79
TW (°C)	26.7	37.8	37.8	37.8	37.8	37.8
OS	1	1	1	1	1	1
DP (μm)	3.5	3.25	3.25	3.25	3.25	3.25
MS	74	74.1	74.1	74.1	74.1	74.1
SG (g/cm^3)	2.24	2.24	2.24	2.24	2.24	2.24
XC (%)	50	50	50	50	50	50
XE (%)	6	6	6	6	6	6
KR (cm^3/mole–s)	1 x 10^7	0	0	0	0	0

*Values reported are the average values for the data set.

[a] Parameter symbols are defined in Table 1.

[b] Utilized the pulse jet baghouse

[c] Utilized the shaker baghouse

TABLE 3. CONVERSION FACTORS TO SI UNITS

Mass:	10^{-3} kg/g
Length:	0.0245 m/in. 0.3048 m/ft
Volumetric Flowrate:	4.72×10^{-4} (m^3/s)/cfm 6.308×10^{-5} (m^3/s)/gpm

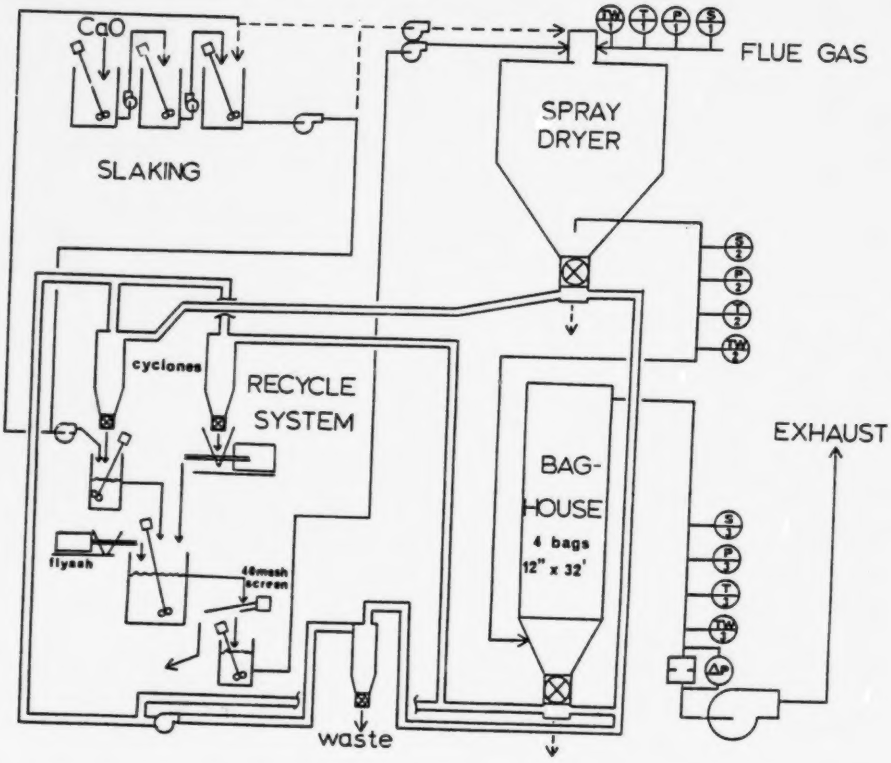

CaO
SLAKING
SPRAY DRYER
TW
T
P
S
FLUE GAS
S 2
P 2
T 2
TW 2
cyclones
RECYCLE SYSTEM
EXHAUST
BAG-HOUSE
4 bags
12" x 32'
flyash
48mesh screen
S 3
P 3
T 3
TW 3
ΔP
waste

Figure 1. Schematic diagram of the pilot spray dryer/baghouse facility.

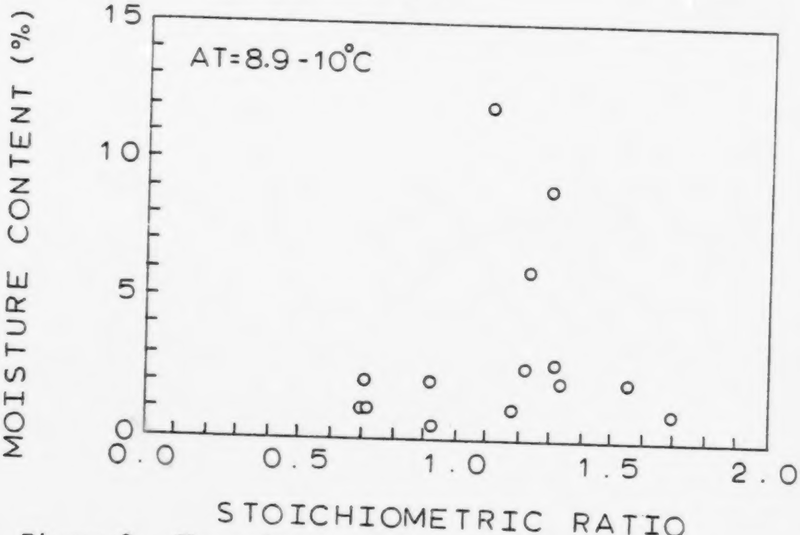


Figure 2. The moisture content of the product as a function of stoichiometric ratio for a $Ca(OH)_2$ slurry (Data Set 4).

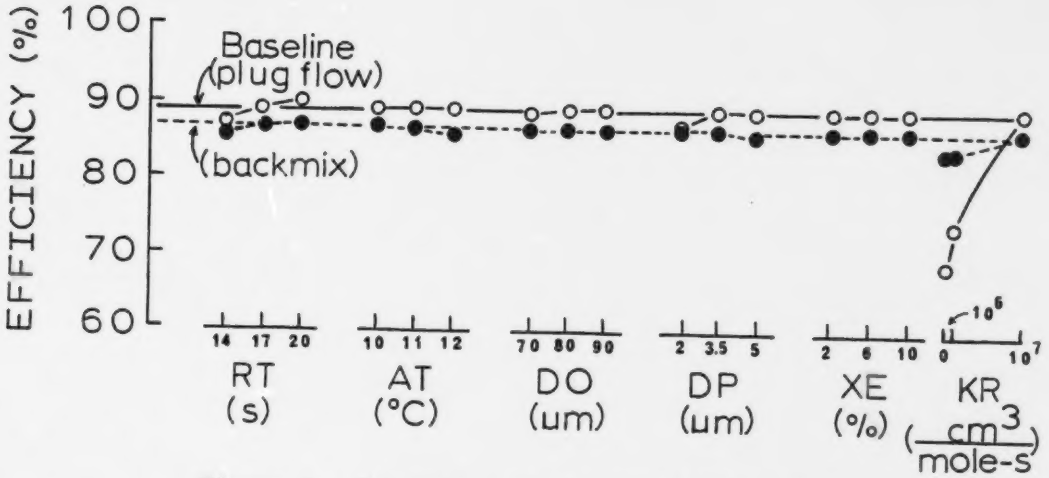


Figure 3. Comparison of the sensitivity of the complete (back) mix and plug flow models.

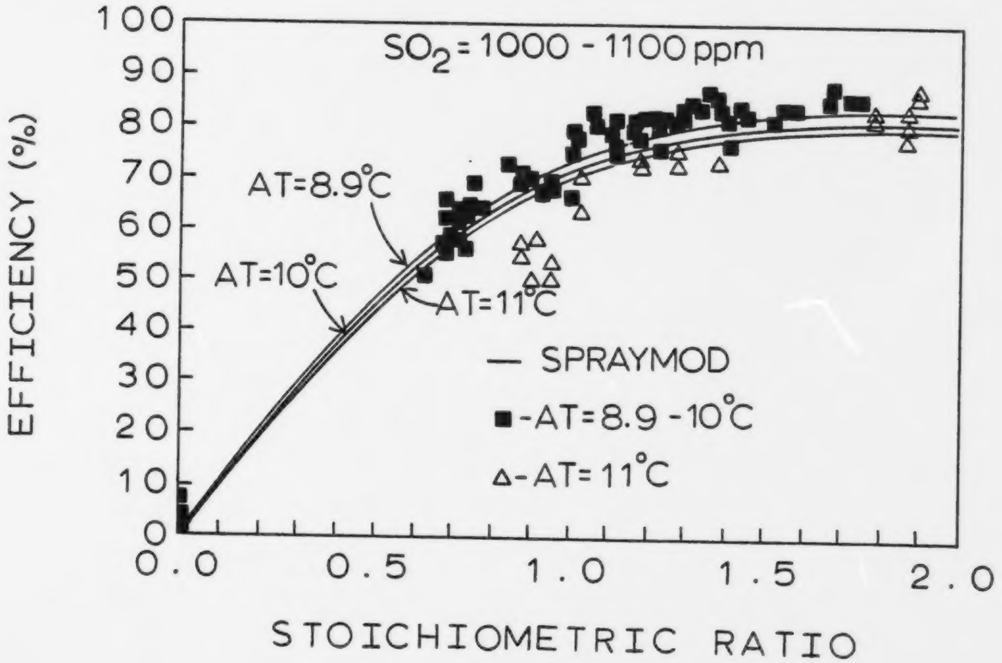


Figure 4. Comparison of SPRAYMOD predicted efficiencies to measured efficiencies for 1000-1100 ppm of SO_2.

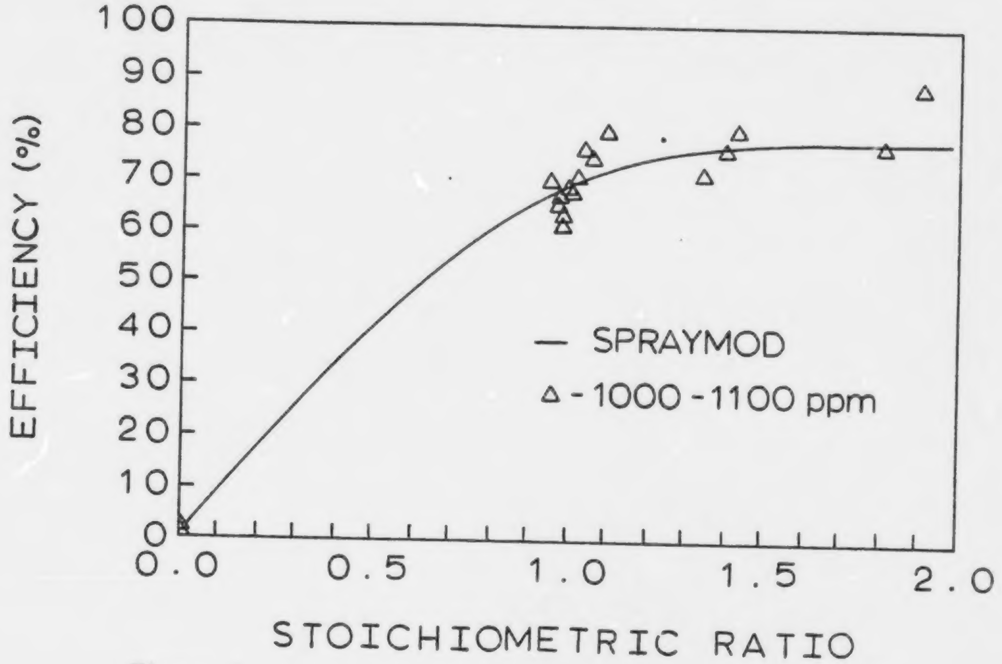


Figure 5. Comparison of SPRAYMOD predicted efficiencies to measured efficiencies for 1000-1100 ppm of SO_2 at a 19.4°C (35°F) approach to saturation.

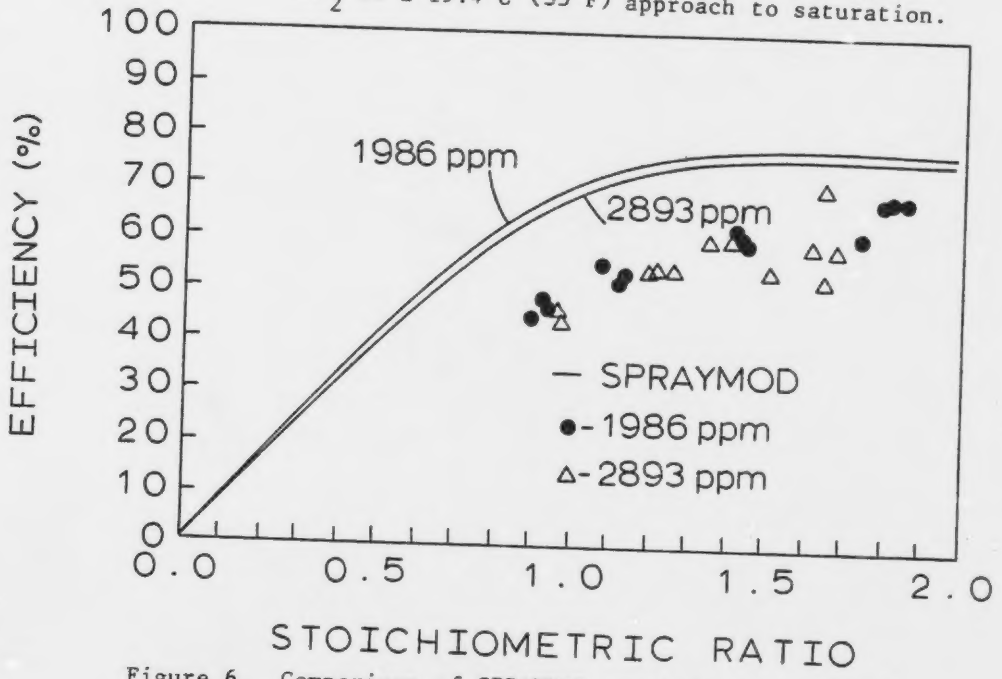


Figure 6. Comparison of SPRAYMOD predicted efficiencies to measured efficiencies for high SO_2 concentrations at 19.4°C (35°F) approach to saturation.

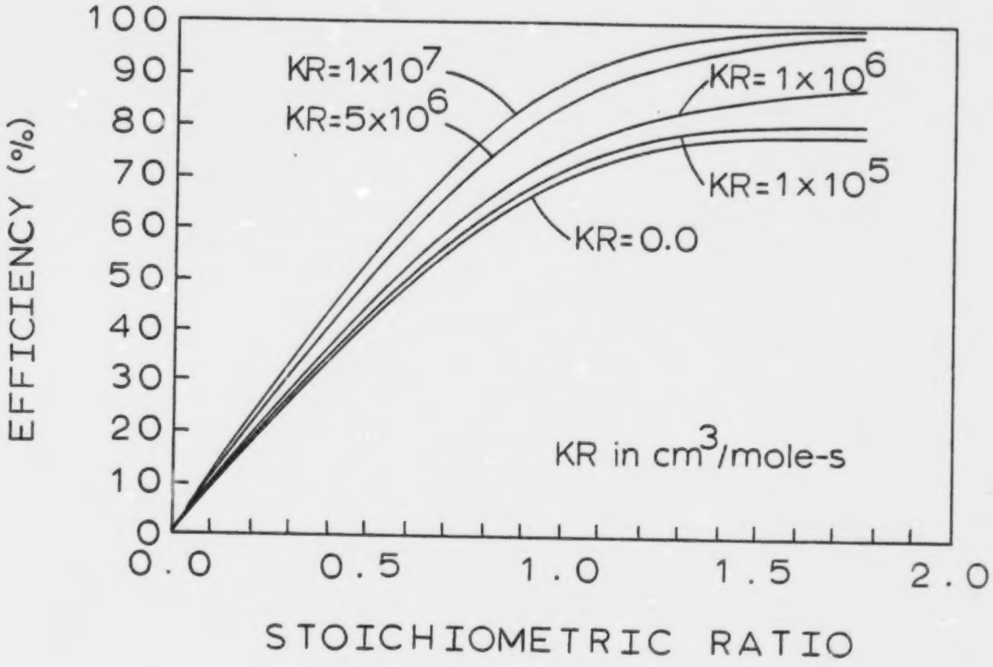


Figure 7. Effect of KR on the efficiency for the baseline condition (plug flow).

Session 10: FGD BY-PRODUCT DISPOSAL/UTILIZATION

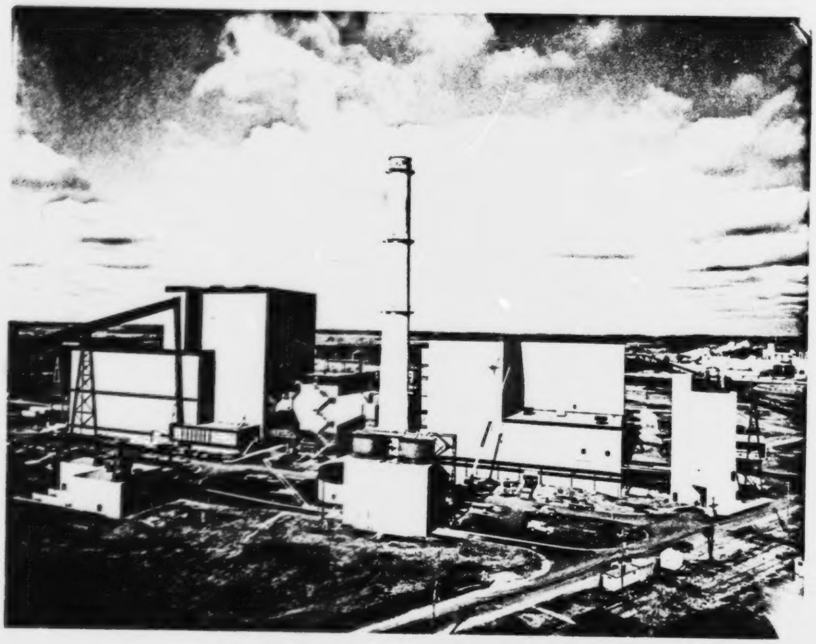
Julian W. Jones, Chairman
U.S. Environmental Protection Agency
Research Triangle Park, NC

GYPSUM BY-PRODUCT FGD SYSTEM

by: W. A. Liegois, P.E.
D. A. Wicks
Stanley Consultants, Inc.
Muscatine, Iowa 52761

ABSTRACT

This paper summarizes the start-up experiences of the FGD system on Unit 9 of Muscatine Power and Water. This FGD system has a guaranteed removal efficiency of 96% of the sulfur dioxide from the flue gas and was the first unit in the U.S. designed to produce a gypsum by-product. Operating problems associated with the system are discussed, along with system availability and by-product quality.



Reproduced from best available copy.

INTRODUCTION

Muscatine Power & Water (MP&W) is a municipal utility serving the city of Muscatine, Iowa. The service area includes light and heavy industry, commercial businesses, and the community of approximately 25,000. Located on the Mississippi River, MP&W has access to ample water supply and barge facilities to accommodate coal and limestone deliveries. In 1978, the utility was operating four small coal-fired steam electric generating plants with a total capacity of approximately 120 MW. The largest of these units was a 66 MW cyclone-fired unit. At that time, Stanley Consultants began the design of Unit 9, a new 160 MW pulverized coal-fired facility. When completed, Unit 9 would more than double the generating capacity of MP&W. In addition, Unit 9 would be the first steam electric generating plant in Iowa to install a flue gas desulfurization system for the control of sulfur dioxide emissions.

MP&W is committed to meeting all applicable local, state, and federal regulations. In addition, the unit has to achieve a high degree of reliability because it represents such a major portion of MP&W's generating capacity. Unit reliability, availability, and maintainability were therefore given high priority in the design of the pollution control equipment. Illinois coal, having a sulfur concentration of approximately 3.5%, will be burned in Unit 9. Based on the sulfur content of this coal and the New Source Performance Standards, a removal efficiency of 90% would be required. Iowa State permits, however, required a removal efficiency of 92%. This is due to modeling by EPA showing Muscatine to be a nonattainment area for SO_2. EPA NSPS therefore requires LAER and SO_2 offsets.

Disposal of fly ash and scrubber-generated waste was also a major concern. On-site ponding of scrubber waste was examined along with several landfill options, including local disposal and transport back to the mine. Potential markets for fly ash and scrubber waste were also explored. Although definite users could not be identified for either the fly ash or the scrubber waste, potential markets do exist. The scrubber waste market is centered around producing an oxidized material, which is high enough in calcium sulfate to replace the naturally occurring mineral, gypsum which is mined in the area. The production of an oxidized scrubber waste has several advantages:

- Improved dewatering characteristics
- Not thixotropic
- Does not require set-up time
- Potential by-product sale (gypsum)

MP&W and SCI believed that an FGD system capable of producing an oxidized waste product would be beneficial. Unit 9 is the first U.S. FGD system design to produce a high grade gypsum material rather than an oxidized waste product.

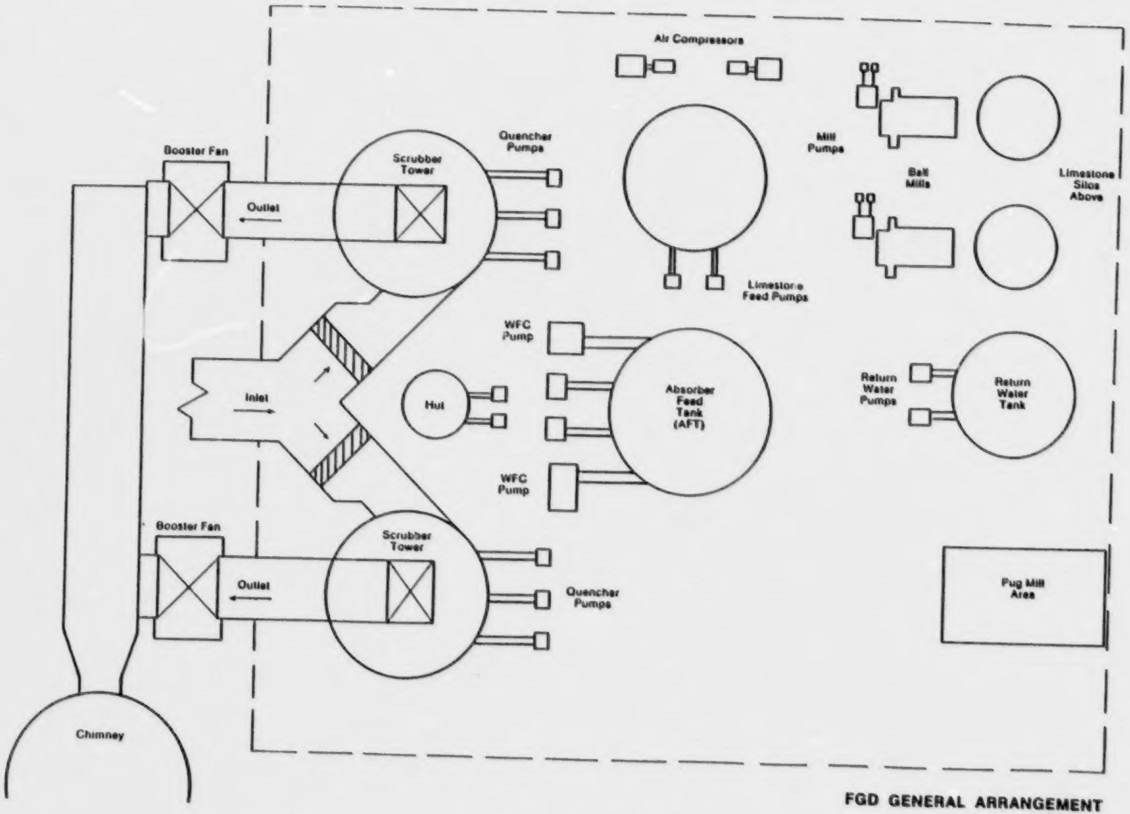
SYSTEM DESIGN

Competitive bids were taken on a complete flue gas cleaning system with a single supplier responsible for both particulate and sulfur dioxide control. Evaluation factors for electricity, water, lime or limestone and steam were included in the specification. All bids received were for wet limestone systems; with Research-Cottrell chosen as the system supplier.

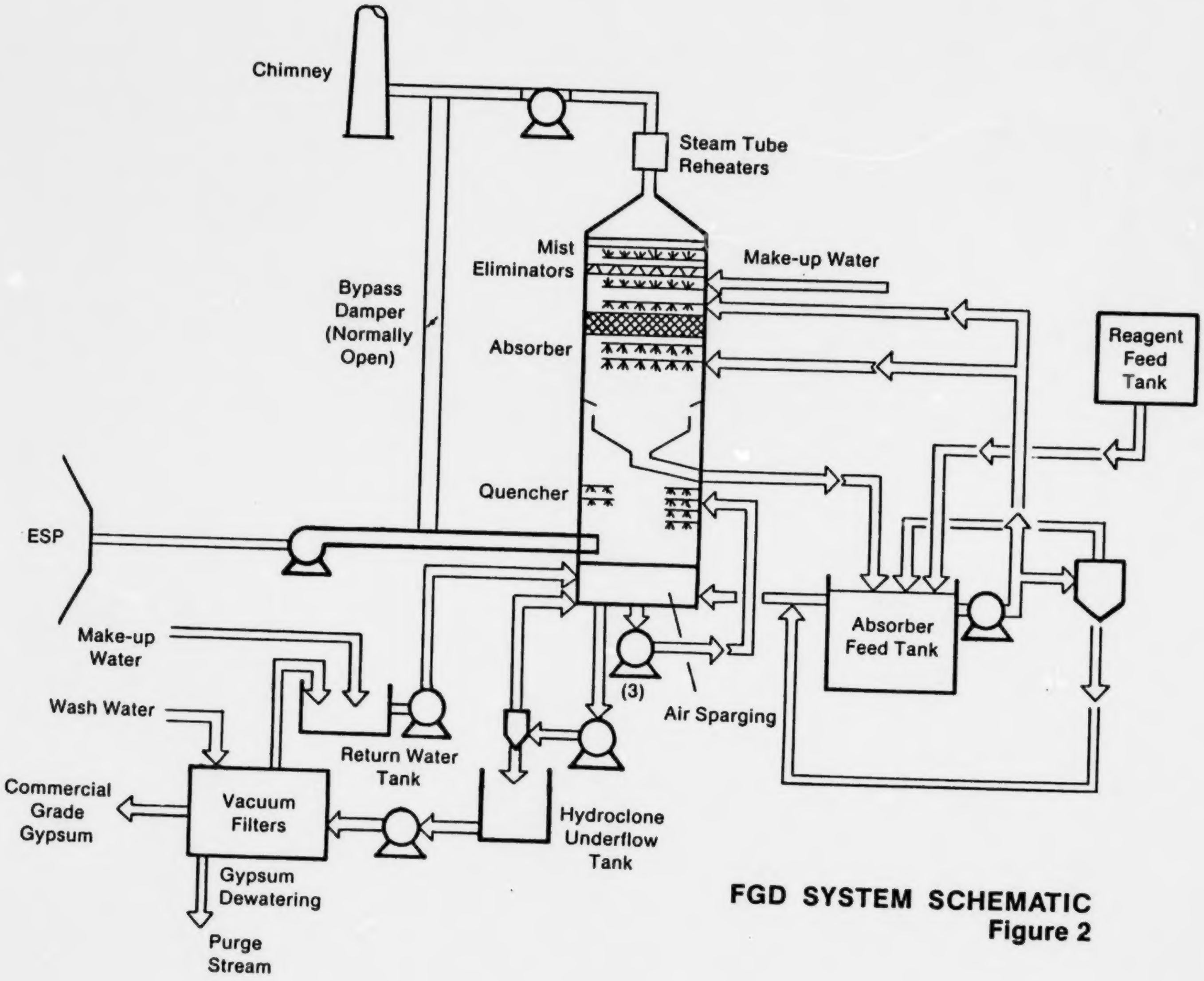
Two electrostatic precipitators (ESP) are used to remove fly ash from the flue gas before it enters the FGD system. Each precipitator is designed to handle 50% of the flue gas and achieve a removal efficiency of 99.7% which represents an emission rate of 0.03 lb of fly ash per million Btu. The ESPs are located on the cold side of the air heater and use rigid discharge electrodes. A crossover duct is located on the inlet of each ESP, and one I.D. fan is located after each ESP. This allows the unit to operate at partial load with any combination of ESPs or I.D. fans in use.

The outlet of the I.D. fans is combined into a single duct which supplies one of two 100% capacity wet scrubbers (see Figure 1). A booster fan is located on the outlet of each scrubber, resulting in the towers operating under a slight vacuum. This system was selected [illegible]use the scrubbers are completely enclosed, and leakage of flue gas into the building was a concern. Finally, a bypass duct from the I.D. fan outlet to the booster fan outlet was included in the design. This duct is equipped with a damper which is normally open. It is closed only when the unit is down and maintenance is being performed. The speed of the booster fans is controlled to maintain a condition of essentially no-flow in this bypass duct. Operating with this bypass damper open protects the boiler from any pressure surges caused by the FGD system or implosion on loss of an F.D. fan. The I.D. fan alone cannot develop enough head to damage the boiler, but the combined heads of the I.D. fan and booster fan could implode the boiler.

The scrubbing towers operate on a double loop principle (see Figure 2). The flue gas enters the bottom of the vertical spray tower and is contacted with a quench spray of limestone slurry with a pH of approximately 4.5. The bottom of the tower operates as both a sump for the quencher spray pumps and a reaction tank for the oxidation air sparged into the bottom of the tower. Three quencher spray pumps are provided for each scrubbing tower, each having its own dedicated spray header. Two 100% oxidation air compressors supply air to either scrubbing tower. Sixteen hydroclones are used in place of thickeners to remove solids from the quencher section. Each tower has two 100% capacity hydroclone feed pumps. These hydroclones can generate an underflow slurry of 50% solids and can be operated with either tower. Hydroclone overflow is recycled back to the quencher section of the tower in operation. Hydroclone underflow drains by gravity to a holding tank and then is pumped to two 100% capacity vacuum filters which produce a filter cake with a contractually required solids content of 85%. This material can either be mixed with fly ash in two 100% capacity pug mills or directly stacked out for sale or disposal. Water from the vacuum filters is stored in a recycle water tank for use as makeup to the quencher section. The plant is designed as a zero discharge facility.


Air Compressors
Booster Fan
Scrubber Tower
Quencher Pumps
Mill Pumps
Ball Mills
Limestone Silos Above
Outlet
Limestone Feed Pumps
WFC Pump
Inlet
Hut
Absorber Feed Tank (AFT)
Return Water Pumps
Return Water Tank
WFC Pump
Booster Fan
Scrubber Tower
Outlet
Quencher Pumps
Pug Mill Area
Chimney

FGD GENERAL ARRANGEMENT
Figure 1


Chimney
Steam Tube Reheaters
Mist Eliminators
Make-up Water
Bypass Damper (Normally Open)
Absorber
Reagent Feed Tank
Quencher
ESP
Absorber Feed Tank
Make-up Water
(3)
Air Sparging
Wash Water
Return Water Tank
Commercial Grade Gypsum
Vacuum Filters
Hydroclone Underflow Tank
Gypsum Dewatering
Purge Stream

FGD SYSTEM SCHEMATIC
Figure 2

The flue gas, after having contacted with the quencher spray, moves up around the bowl and into the absorber section. In this section, the flue gas is first contacted with a limestone spray having a pH just above 6. The flue gas continues up through the tower into a section of packing called the wetted film contactor (WFC). This packing is continuously sprayed from above with the same limestone slurry used in the open spray portion of the absorber section. This packing removes the last portion of sulfur dioxide from the flue gas. Sulfur dioxide concentrations going into the WFC are kept below 600 ppm, and typical outlet concentrations run between 150 and 200 ppm. While the permitted removal efficiency is 92%, the contractual requirement is 96%. The top of the tower contains a two-stage mist eliminator and steam tube reheaters.

Limestone slurry from the absorber section of the tower is collected in the tower bowl and flows by gravity into the absorber feed tank (AFT). Only one AFT is provided for use with both towers. Absorber spray and WFC pumps recirculate slurry to the absorber section. Sixteen hydroclones are used to remove solids from the AFT. The absorber spray pumps also provide the slurry feed to the absorber hydroclones. Hydroclone overflow is returned to the AFT while underflow goes to the quencher sump. An overflow line from the AFT to the tower quencher sump maintains level in the AFT. Water which leaves the AFT through either the overflow line or hydroclone underflow is made up with the fresh limestone feed or the fresh water used to wash the mist eliminators.

Two 100% ball mills are used to prepare limestone and to provide limestone slurry. Ball mill hydroclones are used to recycle oversized particles back to the ball mill feed for regrinding. A single limestone slurry tank is provided for surge capacity.

In general, all pumps and operating equipment have 100% capacity installed spares. All tanks, on the other hand, supply both towers. It is not possible to operate both scrubbing towers at the same time.

GYPSUM BY-PRODUCT SPECIFICATIONS

The by-product gypsum was guaranteed to meet the specifications given in Table 1. These product specifications were to be met regardless of the type of coal being burned.

TABLE 1. GUARANTEED GYPSUM QUALITY

Calcium Sulfate ($CaSO_4 \cdot 2H_2O$)	94.0% min.
Calcium Sulfite ($CaSO_3 \cdot 1/2\ H_2O$)	0.5% max.
Calcium Carbonate ($CaCO_3$)	1.5% max.
Insolubles	3.5% max.
Total Solids	85.0% min.

Note: Guarantees based on total solids.

The soluble ion concentrations, however, are dependent on the coal burned, type of limestone used, quality of the fresh water used, and the use of a fresh water for vacuum filter cake wash. An FGD system water purge was recommended as another means of controlling the soluble ion concentrations in the scrubber loop and in the by-product gypsum. Table 2 gives anticipated maximum concentrations of soluble ions in the by-product gypsum based on the design coal with two different water purge rates.

TABLE 2. GYPSUM ION CONCENTRATIONS

No Filter Cake Wash 21 gpm Water Purge		With Filter Cake Wash 35 gpm Water Purge
Sodium	150 ppm	75 ppm
Magnesium	500 ppm	175 ppm
Potassium	100 ppm	50 ppm
Chloride	1150 ppm	400 ppm
Other	80 ppm	80 ppm

PROBLEM AREAS

The FGD system at MP&W is very similar to other units the system supplier has installed. It is, however, the first unit designed to produce a by-product gypsum, not just an oxidized waste product. This has resulted in more problems during start-up. The unit must not only remove 96% of the sulfur dioxide from the flue gas but it must also meet by-product specifications. All of this must be accomplished without exceeding the other system guarantees for use of electricity, steam, water, and limestone.

Several specific problem areas have been encountered. The following is a brief description of the major problems.

QUENCHER LINING

The scrubbing tower was originally designed to be an all alloy steel construction. The upper section of the tower, downstream ductwork, and bypass duct are all 316L material. The quencher section of the tower, including the tower sump, however, is 904L. The steam reheaters are Alloy G. The chloride content of some of the coals designed to be burned at Unit 9 are quite high. With an oxidized scrubbing system and dewatering the waste product to 85% solids, not much free water leaves the system, resulting in predicted soluble chloride ion concentration as high as 20,000 ppm. During construction, the system supplier recommended lining the quencher section of the tower.

A spray-on rubber lining was installed in the quencher sump with two rows of acid block at the top just above the high water overflow point. The rubber lining had to be installed twice because of poor adhesion. Maintaining adhesion of the rubber lining on the 904L base metal was very difficult.

Initial operation of the tower showed that hard calcium sulfate scale falling into the quencher sump would damage the acid block. A hard ceramic lip was installed on top of the block in an effort to protect the block. This solution has been only partially successful. Falling scale still damages portions of the block and the ceramic lip but damage is relatively minor and easy to repair. There is not, however, any indication of corrosion in the tower. A thick (1-2 inch) layer of hard scale deposit has built up on much of the quencher section. Blasting this material with a fire hose has little effect. Occasionally, patches of this scale will break off. No attempt has been made to remove any material that does not come off when blasted with the fire hose. A final evaluation of corrosion in this section of the tower will not be made until closed loop operation is achieved and high soluble ion concentrations occur.

AFT DENSITY CONTROL

The FGD system has been able to consistently meet the 92% sulfur dioxide removal efficiency required by state permit. The system design, however, called for approximately half of this removal to be achieved in the quencher section. In actual operation, only 10-30% of the sulfur dioxide is removed in the quencher section; the balance is removed in the absorber section. This means that many more solids must be removed from the AFT by the absorber hydroclones. It also has meant that the solids residence time in the AFT is much shorter than design.

The number of absorber hydroclones has been increased from two to sixteen in an effort to remove the additional solids. This modification has been only partially successful. The major problem is the particle size of the calcium sulfite crystals. Because the absorber operates at a pH just above 6, most of the solids in the AFT are calcium sulfite. At Unit 9, these calcium sulfite crystals are smaller than the unreacted limestone and the calcium sulfate which is formed. The hydroclones, therefore, begin to remove limestone from the AFT before calcium sulfite. More limestone is fed to the AFT to maintain pH, but the extra limestone ends up in the quencher section, raising its pH above 5.

A potential solution to this problem centered on reducing the size of the limestone particles. If they were smaller than the calcium sulfite crystals, the operation of the hydroclones would be improved. The original system design was based on the limestone particles being smaller than both the calcium sulfite and sulfate crystals. The recycle hydroclones on the limestone ball mills were modified to produce a finer limestone slurry. In order to meet the limestone requirements at full load with the modified ball mill hydroclones, both limestone ball mills were required. This did result in some improvement in solids removal but not enough to correct the problem, so the limestone ball mills were returned to their original design.

Recent testing has shown improved operation with the AFT maintained at a pH just below 6. This results in less limestone in the AFT and less limestone removed to the quencher by the absorber hydroclones. Hopefully, this solution will prove successful. In the meantime, the system supplier is investigating other options for AFT solids control.

QUENCHER pH CONTROL

The pH of the quencher section has operated above the design point of 4.5 because of the AFT hydroclones. The extra solids removed from the AFT contain large amounts of limestone. The amount of limestone moved to the quencher section by the absorber hydroclones is much higher than design. As the quencher pH rises, the oxidation efficiency of calcium sulfite to calcium sulfate falls. The gypsum product is therefore high in both limestone and calcium sulfite. This has also had a detrimental effect on the operation of the vacuum filters. Waste product which is high in sulfite and limestone is harder to dewater and can become thixotropic.

To correct this problem, a sulfuric acid addition system has been installed. Sulfuric acid can be added to the quencher to react with the excess limestone and reduce the pH. This system has only been in operation a few weeks and results are very preliminary. The addition did improve the waste product quality but the addition of acid must be carefully controlled to guard against overfeeding, which will reduce the sulfur dioxide removal efficiency of the tower. This acid addition system will not be needed if operation of the AFT at a reduced pH (5.8 to 6.0) produces all of the desired results. The problem of the quencher pH cannot be separated from the AFT density control problem.

BYPASS FLOW

The open bypass was designed to protect the boiler from pressure surges caused by the FGD system and to protect the boiler from the combined pressure differential the ID and booster fans can develop. Booster fan speed is used to control the amount of flue gas fed to the FGD system. If the booster fan is run too slowly, flue gas will bypass the scrubber. The bypassing of any flue gas will significantly reduce the system sulfur dioxide removal efficiency when trying to maintain a permitted level of 92% and a guaranteed removal efficiency of 96%. If the booster fan is run too fast, scrubbed flue gas will be recycled. If recycling occurs at full load, the gas velocity in the scrubber will be above design. This can reduce the scrubbing efficiency and cause fouling in the tower. Recycling gas at lower boiler loads can also cause unnecessary fouling.

Initially, the speed of the booster fan was set by the boiler load. The pressure drop across the scrubbing tower, however, was not constant due to fouling of the wetted film contactor. A correlation has been developed between gas flow and the pressure drop between the scrubber inlet and just above the bowl in the absorber section of the tower. This pressure drop over the bottom portion of the tower is not affected by fouling. The increased pressure drop caused by fouling occurs across the WFC. The boiler load is now used to determine the gas flow through the scrubber, and pressure drop across the bottom of the tower will determine the booster fan speed. Two flow sensors have also been placed in the bypass duct. Each sensor measures flow in either direction. Alarm lights are located on the scrubber control panel. The sensitivity of the switches is set so there is a small dead band representing an essentially no-flow condition. A white light is lit if neither sensor shows a flow. Four red lights are used to indicate either bypass or recycle flow for both sensors.

Both the flow indicators and the revised booster fan operating philosophy have just been installed. They should improve the booster fan operation and reduce the amount of recycling or bypassing.

PACKING PROBLEMS

While there has been some buildup of solids on the walls of the tower and spray headers, the biggest problem area has been with the packing in the wetted film contractor. There have probably been several reasons for the fouling experienced in this part of the tower, and at different times each may have resulted in the packing becoming plugged. The most likely causes have been:

- High gas velocity due to higher than design gas volume in the scrubber
- High solids content in the absorber spray slurry
- High limestone concentrations in the absorber spray slurry
- High sulfur dioxide concentrations going into the packing

Steps have been taken to control the gas flow to the scrubber which should prevent this problem from recurring. The improved booster fan controls will also help. The chemistry of the AFT is being watched more closely than before in an effort to control both the total solids and the limestone concentration. Operation of the AFT at the lower pH should reduce fouling. Tests have also just been completed to determine the concentration of sulfur dioxide in the flue gas at different points in the scrubber. Overworking the packing by allowing high concentrations of sulfur dioxide to enter the wetted film contractor can cause pluggage because of the additional solids generated. These tests show that the concentration of sulfur dioxide entering the WFC can be kept below the design value of 600 ppm.

Once fouling begins, the situation is irreversible because the pluggage reduces the area for gas to pass through. As the same volume of gas passes through the reduced area, the velocity increases. This prevents proper drainage and further floods the packing. A tower which has operated for weeks or even months without any fouling can become almost totally plugged in a matter of days.

Cleaning of the plugged packing has become a second problem. The packing in use at MP&W is a thin-wall corrugated polypropylene type. This packing has demonstrated excellent sulfur dioxide removal capability. It is, however, difficult to clean and can be easily broken during cleaning. The system supplier has been testing several different types and manufacturers in an effort to find a more durable and more easily cleaned packing. This must be done while still providing sufficient contacting to achieve the required sulfur dioxide removal efficiency. The testing has included several new packing designs of his own. At the current time, MP&W is still using the original packing design. It is hoped that a new packing design which is made of a rigid PVC and is more open for easier cleaning will be available for full-scale testing this fall at MP&W.

CLOSED LOOP OPERATION

The system was designed to operate without any liquid purge or blowdown. A blowdown might be necessary when trying to make a high purity gypsum product while burning high chloride coal. Under normal operations, however, a purge would not be necessary. Fresh water is designed to come into the system in four ways:

- Through pump and agitator seals
- With the fresh limestone feed
- From the mist eliminator wash, which always uses fresh water
- As makeup to the return water tank.

Because the waste product gypsum is so highly dewatered by the vacuum filters, very little fresh water is necessary. Close attention must be paid to each of the fresh water sources if the water balance is to be maintained.

The water balance is complicated because the mist eliminator wash cycle and the limestone feed are not continuous suppliers of water. Also, the vacuum filter does not operate continuously. Vacuum filters work best when operated at full load. At MP&W, two 100% capacity filters were provided. This means that one filter is cycling on and off at a rate depending on boiler load. This results in large surges in the return water tank level. Fresh make-up water is added to this tank on low level, and high tank level is controlled by an overflow to the floor drains. The capacity of this tank was almost doubled from the original design to allow the level to swing over a wider range before overflowing. It is hoped that this modification will allow Unit 9 to operate closed loop.

It is important that water balances be done at various boiler loads. The amount of water added to the system from seals and the mist eliminator wash does not usually vary with load. At low loads, the amount of water lost through evaporation and with the waste product is greatly reduced. This can result in a system that goes into a net positive water balance at reduced boiler load. To correct this problem: reduce the amount of seal water used by either shutting off unnecessary pumps and seals or by installing low water use seals; secondly, the mist eliminator wash cycle frequency can be lengthened at low loads, reducing the amount of water used. Low load operation is the most difficult water balance condition.

FLOOR TRENCHES

The scrubbing system for Unit 9 is totally enclosed in a building. Floor trenches were provided along with four sump pumps at two different locations. The sump pumps transport the wastewater to the coal pile run-off pond prior to being processed in the water treatment plant. Rainwater and any drainage from the stackout and loading area is also collected in a sump and pumped to the coal pile runoff pond.

Two problems became obvious very quickly. First, even though the scrubbers and their associated tanks were not scheduled to be cleaned very often, during initial start-up and shakedown this ocurred more than anticipated. This caused the coal pile runoff pond to quickly begin to fill with scrubber sludge. A new, two-cell settling basin was built to receive the wastewaters from all the FGD sumps. The liquid overflow from this basin goes to the coal pile runoff pond and then on to the water treatment plant. When one cell becomes full, feed is shifted to the other and underflow drains are opened. The basin is designed with an inclined ramp at one end to allow a front end loader to be driven in to clean out the cell. The two cells are continuously rotated.

The second problem centered around the floor trenches themselves. Waste material (especially oxidized scrubber waste) drained into the trenches would quickly settle out unless water was flushed down the trench to keep it moving. The floor trenches were much easier to clean than underfloor drains. Hose stations, however, should be located near equipment that will be drained. These hoses can be used to flush the slurry through the trenches. All equipment should be located on raised pads in case a floor trench should become plugged and flood the area.

POSITIVE SYSTEM ASPECTS

Several positive aspects of the system have also been noticed. Listed are a few of the major items.

REMOVAL EFFICIENCY

The system has been able to meet the permitted level of 92% removal. It normally operates at 94% removal or better. Hopefully the work currently underway will enable the system to achieve the 96% removal guaranteed by the system supplier.

FILTER CAKE

The system produces a mostly oxidized waste product which is normally easy to dewater and handle. The waste can either be stacked out or it can be blended with fly ash. In either case, it can be handled immediately.

CORROSION

The system has shown no evidence of corrosion. All vessels, tanks, and ductwork appear in good condition.

AVAILABILITY

System availability, operability, and reliability figures from January 1984 to April 1985 are given in Table 3. The months of March and April 1984 and April 1985 were scheduled boiler outages. Unscheduled boiler outages occurred in July and November 1984. On only one occasion has boiler operation been affected by the FGD system. Two absorber spray pumps are required for full load operation, and vibration in one of the absorber spray pumps required the boiler to operate at half load for two to three days while repairs were completed.

The use of spare pumps, ball mills, towers, vacuum filters, and pug mills has allowed many system changes and modifications to be completed while the unit was in operation. Fortunately, these system changes and modifications have never kept the backup equipment from being available when needed. System operability and reliability have been excellant. The unit has remained online and has consistently met the 92% removal requirements for the control of sulfur dioxide.

TABLE 3. SYSTEM AVAILABILITY, OPERABILITY, RELIABILITY

	1984							
	J	F	M	A	M	J	J	A
Availability	72.11	100	0	24.9	100	100	68	90.3
Operability	48.75	95	0	60.2	98.2	100	97.6	98.6
Reliability	100	100	0	100	100	100	100	100

	1984				1985			
	S	O	N	D	J	F	M	A
Availability	100	84	6.7	99.9	100	98.4	100	14.4
Operability	99.6	100	68.2	98.35	95.2	98.6	98.9	86.2
Reliability	100	100	100	99.9	100	97.5	98.85	100

GYPSUM PRODUCT QUALITY

The guaranteed gypsum product specifications given in Table 1 are checked daily. Because of the problem areas discussed, the gypsum product does not normally meet these specifications. The high limestone content in the quencher section results in excessive amounts of limestone in the gypsum product. Limestone concentrations routinely run around 2% - 10%. The oxidation efficiency in the quencher has also been reduced because of the high pH caused by the excess limestone. Calcium sulfite concentrations of 0.5% to 4.0% are common. Finally, because of the high limestone and calcium sulfite concentrations in the gypsum, the vacuum filters normally produce a cake containing 75% to 80% solids.

Past testing has shown gypsum product specifications can be met if the AFT density control problems are corrected. If the modifications completed during the April 1985 outage are successful in controlling the AFT density, then the gypsum product quality should improve significantly.

SUMMARY

In conclusion, the system has met most of its major objectives. System operability and reliability have been good. The sulfur dioxide removal efficiency has been good enough to meet the 92% removal required by permits. Removal efficiencies of 94%-95% are common and hopefully the system revisions currently being tested will produce efficiencies of 96%. While gypsum product quality has not achieved design levels, the waste material is easy to handle and dewater, and blends well with fly ash. The unit is currently operating and scheduled for performance testing during June and July 1985.

The authors would like to acknowledge the help and cooperation of the staff of Muscatine Power and Water. Special thanks go to Ray Donz, Tom Cox, and Jerry Millage.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

BY-PRODUCT GYPSUM PRODUCTION AT A 2300 MW POWER PLANT

by: Don Mzyk
Texas Utilities Generating Co.
Martin Lake Station
Tatum, Texas 75691

Jan Zmuda
Research-Cottrell, Inc.
Maintenance & Service Division
Somerville, NJ 08876

ABSTRACT

The Martin Lake Station of Texas Utilities consists of three 770 MW boilers firing lignite coal. Each boiler is equipped with a Research-Cottrell limestone flue gas desulfurization (FGD) system. The FGD systems were originally designed to produce a sulfite sludge to be blended with fly ash for disposal.

In early 1983 each FGD system was retrofitted with equipment for in-situ oxidation to produce a calcium sulfate (gypsum) by-product. This paper will discuss startup and testing of the forced oxidation system. Utility operating experience, which led to the signing of a contract for the sale of commercial grade gypsum for wallboard manufacture, will also be discussed.

INTRODUCTION

Texas Utilities Generating Company has been operating flue gas desulfurization (FGD) systems at Martin Lake Station for over eight years. These systems were the first utility application of FGD on lignite-fired boilers.

Texas Utilities has been very progressive in its activities to continually improve the operation of their plants. In 1980, Research-Cottrell was issued a contract to upgrade the FGD system to permit burning of higher sulfer lignite. This upgrade included in-situ forced oxidation to produce a calcium sulfate (gypsum) by-product. During startup in mid-1983, it became apparent that commercial grade gypsum could be produced consistently.

Texas Utilities actively began to pursue the approach of producing a commercial grade by-product. A tighter control of process operation and maintenance, coupled with an extensive analytical program, was initiated. These activities led to the signing of a contract in late 1984 for the sale of a wallboard-grade gypsum by-product. Texas Utilities will become the first utility operating a limestone-based FGD system to produce and sell a commercial grade gypsum in the United States. This paper will discuss these activities.

Description of Martin Lake Station

The Martin Lake Steam Electric Station of Texas Utilities Generating Company is located in Rusk County, near Tatum, Texas. Each of the three lignite-fired units can produce over 770 MW at maximum continuous rating, for a total of more than 2300 MW in gross generation.

Lignite, which is mined within ten miles of the station, is fired at a rate greater than 14,000 tons/day, per boiler, at maximum continuous rating. Table 1 shows the typical composition of the lignite fuel.

TABLE 1. TYPICAL COMPOSITION OF LIGNITE FUEL

Specification	Range
Heating Value, Btu/Lb	6900 - 7900 (Original) 6000 - 7200 (Revised)
Ash Content, %	8 - 13
Moisture Content, %	29.0 - 37.9
Sulfer Content, %	0.5 - 1.5 (Original) 0.9 - 3.8 (Revised)

Each boiler is equipped with a Research-Cottrell electrostatic precipitator for control of particulate. Each precipitator provides greater than 99.4% removal efficiency, to meet a 0.1 lbs/mmBTU emissions code. The SO_2 emissions regulation for this plant is 1.2 lbs/mmBTU. To comply with this regulation, each unit was supplied with a Research-Cottrell Double-Loop™ flue gas desulfurization system.

The FGD system was originally designed to treat up to 75% of the flue gas generated in the boiler. The remaining flue gas was by-passed to provide reheat for the treated flue gas. Units 1, 2 and 3 went on-line in April, 1977; May, 1978; and February, 1979, respectively. In 1980, Research-Cottrell was contracted to retrofit additional scrubber capacity because of an anticipated increase in the sulfur content of the lignite. Two additional towers were added, for a total of eight towers per unit. This equipment was started up in 1983.

A brief description of the FGD system as supplied (and upgraded) is provided.

FGD System Description

Figure 1 shows the general arrangement of the absorber towers at the Martin Lake Station. The eight absorber towers are grouped into four pairs of towers. Each tower pair, or module, is fed by a common absorber feed tank (recycle tank). Flue gas at 3,900,000 acfm and 360°F enters the inlet manifold to the scrubber towers. Based on the inlet sulfur level, from 75% to 95% of the flue gas is treated. The remaining 5% to 25% of the flue gas by-passes the system to the outlet duct.

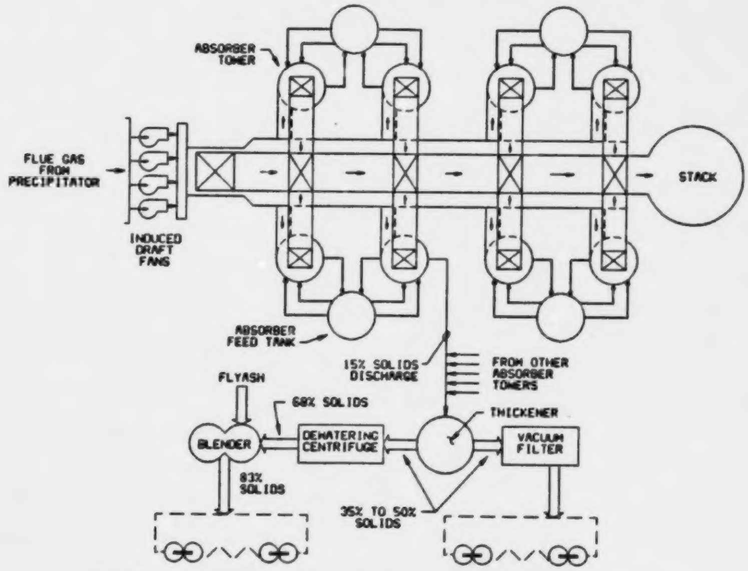

ABSORBER TOWER
FLUE GAS FROM PRECIPITATOR
STACK
INDUCED DRAFT FANS
ABSORBER FEED TANK
15% SOLIDS DISCHARGE
FROM OTHER ABSORBER TOWERS
FLYASH
68% SOLIDS
THICKENER
BLENDER
DEWATERING CENTRIFUGE
VACUUM FILTER
83% SOLIDS
35% TO 50% SOLIDS

FIGURE 1 - GENERAL ARRANGEMENT OF THE MARTIN LAKE FGD SYSTEM

A schematic of a typical absorber tower is shown in Figure 2. Flue gas enters the cyclonic quencher where it is contacted by slurry and cooled to a temperature of about 140°F. The quencher also removes up to 35% of the entering SO_2 as well as virtually all of any remaining particulates entering the FGD.

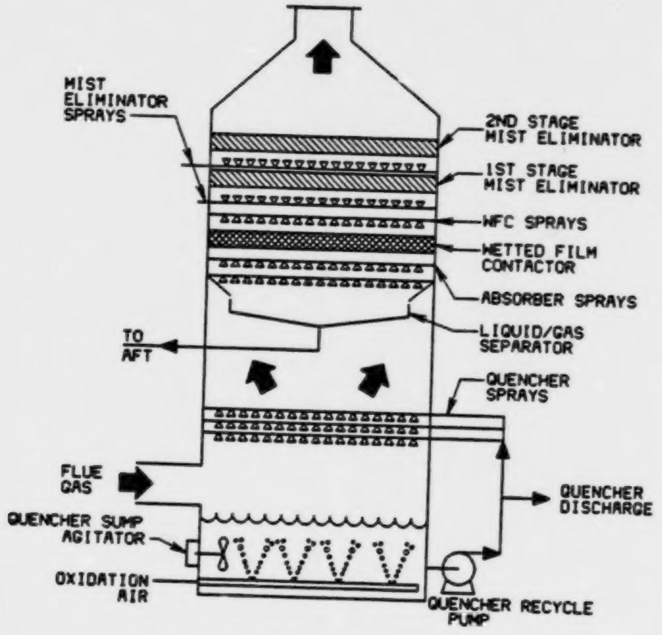

MIST ELIMINATOR SPRAYS
2ND STAGE MIST ELIMINATOR
1ST STAGE MIST ELIMINATOR
WFC SPRAYS
WETTED FILM CONTACTOR
ABSORBER SPRAYS
LIQUID/GAS SEPARATOR
TO AFT
QUENCHER SPRAYS
FLUE GAS
QUENCHER DISCHARGE
QUENCHER SUMP AGITATOR
OXIDATION AIR
QUENCHER RECYCLE PUMP

FIGURE 2 - DOUBLE-LOOP™ ABSORBER TOWER

Quenched flue gas then passes through the liquid-gas separator (commonly referred to as the "bowl") into the absorber section. The bowl separates the absorber loop slurry from that of the quencher loop. In the absorber section, the flue gas passes through absorber sprays, wetted film contactor (packing), and a set of mist eliminators. The absorber sprays and tower packing remove a high percentage of the SO_2 entering this section. The mist eliminators remove entrained moisture from the flue gas prior to its entering the outlet duct feeding the stack. Total SO_2 removal in the absorber towers is typically better than 95%.

The treated flue gas from the towers is mixed with any by-passed flue gas in the outlet duct. Additional reheat is provided by heated ambient air.

Limestone is brought to the site in pebble form via rail cars and is ground in wet ball mills. The resulting slurry is stored in a reagent feed tank at a concentration of approximately 40% solids. Reagent is pumped to the FGD systems via a recirculating loop.

Fresh limestone slurry enters the absorber loop through the absorber feed tank. Slurry in this tank consists of fresh reagent and the reaction products, calcium sulfite and calcium sulfate. The slurry is pumped to the spray tower and wetted film contactor and returns to the tank through the bowl return line. Reaction products formed in the tank are discharged by overflow into the quencher. Gypsum seed crystals are recycled from the quencher back into this section to control scaling.

In the quencher section, the partially reacted slurry is circulated through the quencher sprays. This slurry contains primarily reaction products (calcium sulfite and calcium sulfate) plus a small amount of unreacted limestone. Fresh limestone is added to this section to reduce absorber loop SO_2 loading and to provide additional pH control. The purpose of this lower section is to maximize limestone utilization and to provide a reaction vessel for in-situ oxidation. With oxidation, the primary reaction product is calcium sulfate (gypsum).

Quencher slurry containing 15% solids is discharged to the waste disposal system. This system consists of thickeners for primary dewatering followed by secondary dewatering. Overflow from the thickeners is returned to the FGD system; underflow at 30% to 50% solids is pumped to the secondary dewatering building.

Secondary dewatering is accomplished using either centrifuges or vacuum filters. Prior to the 1983 upgrade, all of the sludge was dewatered in centrifuges to about 60% solids. This sludge was then blended with fly ash to produce a dry, non-thixotropic blend which was dumped into rail cars and transported to the disposal site. As part of the upgrade, vacuum filters were added to provide additional dewatering. Thickener underflow is pumped to the vacuum filters and dewatered to greater than 80% solids. The "dry" cake from the vacuum filters is discharged into rail cars. Water recovered from secondary dewatering is recycled back to the FGD system.

TESTING OF FORCED OXIDATION

Prior to the installation and start up of the oxidation equipment (in 1983), Research-Cottrell conducted several oxidation-related test programs at Martin Lake. These programs were performed with the co-operation and assistance of Martin Lake plant personnel.

In 1978, Research-Cottrell modified one tower on Unit 1 to produce a gypsum by-product. One quencher was fitted with an oxidation sparger. Discharge from this quencher was pumped to a hydroclone, which concentrated the slurry to about 35% solids. The concentrated slurry was stored in a holding tank, then pumped to a small rotary drum vacuum filter for dewatering. Dewatered gypsum cake at about 90% solids was discharged into 55 gallon drums. More than 100 metric tons of gypsum were produced. This material was sent to a major European gypsum company for testing and was found to be suitable for use in cement and wallboard manufacture. A large quantity of wallboard sheets were manufactured from this gypsum. Table 2 shows the typical analysis of the material produced during this demonstration program.

Table 2. Chemical Specifications of Martin Lake Gypsum

(1978 Demonstration)

Specifications	Gypsum for Wallboard	Martin Lake Gypsum (Avg)
Moisture of raw gypsum,%	<10	
Degree of purity, %	>94.6	98.8
MgO, ppm	<200	200
Na_2O, ppm	<400	200
Chloride (Cl), ppm	<100	25
Sulfur dioxide (SO_2), ppm	<3000	1000
pH	>5.0 to <7.0	6.5
Potassium Oxide (K_2O), ppm	<400	100
Phosphorous-pentoxide (P_2O_5), ppm	<200	100
Fluoride, ppm	<500	100
Organic matter, ppm	<1000	370
Color, degree of whiteness, %	>80	82

Except for moisure content, all values are referred to dried calcium sulfate-dihydrate at 40° Celsius.

In late 1980 and early 1981, Research-Cottrell was again involved with testing activities at Martin Lake related to the oxidation system. In this program the oxidation equipment planned as part of the high sulfur upgrade was incorporated into one tower. The goals of this program were to evaluate the sparger design, demonstrate the operation of the oxidation system, evaluate the overall performance of a scrubber tower with the oxidation system in service, and to identify any potential problem areas. In conjunction with this program, the filtering and settling characteristics of the oxidized slurry were examined. Filter leaf tests were conducted to simulate various cloth materials, suction vacuums and drum speeds. This data was used as the basis for the rotary drum vacuum filter sizing. One tower on Unit 3 was selected for testing.

During the period from August 1981 to February 1982, additional testing was done at Martin Lake. These tests were a continuation of testing activities conducted earlier in 1981. One tower on Unit 3 was again selected for testing, incorporating improvements defined from previous testing.

MARTIN LAKE FORCED OXIDATION UPGRADE

Because of an anticipated increase in coal sulfur loading. Texas Utilities contracted Research-Cottrell to supply additional scrubber capacity. Studies were conducted of all the various system components-gas handling, reagent prep., dewatering, etc.-to determine the impact of the higher sulfur fuel. The intent was to provide additional scrubbing capacity while maximizing the use of existing equipment.

In-situ forced oxidation was selected as the means to reduce sludge dewatering requirements and provide operating flexibility. With the anticipated increase in sludge volume, additional thickener capacity would have been required for unoxidized sludge. Forced oxidation eliminated the need for additional thickeners, even at the increased sludge volume, resulting in a significant capital savings.

Modifications to the existing thickeners, designed for unoxidized sludge, were required due to the settling characteristics of the gypsum crystals. Gypsum crystals settle much faster than calcium sulfite, causing a rapid buildup of solids, making the existing thickeners too large. A recycle of thickener underflow back into the thickener was incorporated to prevent the solids from concentrating too rapidly. Underflow solids levels above 50% would impact pump operation, plug pipelines, and lead to thickener overloading.

Forced oxidation permitted the use of vacuum filters for secondary dewatering. A dry, non-thixotropic cake could be readily produced with vacuum filters. Therefore, the need for blending with fly ash was eliminated. Separate discharge chutes and railroad tracks for the gypsum cake were incorporated into the upgrade. These permit segregation of the vacuum filter cake if alternates to landfill disposal could be found for the high quality gypsum by-product.

Process and equipment modifications for conversion to forced oxidation sludge treatment included:

- oxidation air compressors
- sparger piping in all quencher sumps
- larger diameter thickener underflow lines
- modifications to thickener underflow pumps
- modifications to existing centrifuges
- rotary drum vacuum filters
- discharge chutes, conveyors, and railroad sidings

In conjunction with the upgrade of the FGD system, a waste water treatment plant was built to bring the Martin Lake FGD System to a "zero discharge" status. This plant consists of three parallel treatment stages for processing wastewater. These include reverse osmosis, chemical softening and evaporation. A purge stream of process water from the FGD system is treated and returned back to the FGD system. This eliminates the need to use fresh water in areas such as pump seal water, demister wash water, and reagent preparation water. Additional benefits of the wastewater treatment include reduced corrosion in the FGD system and the ability to control the concentrations of elements detrimental to the use of the by-product gypsum for wallboard. Distillate from the evaporator can be used for boiler makeup.

Construction of the high sulfur upgrade modifications was completed in early 1983. A stepwise program was undertaken for the startup of the forced oxidation system. This gradual phase-in of forced oxidation was done to:

- familiarize plant personnel to the operation of the upgraded FGD system,
- identify any potential operational changes required in the full scale system,
- optimize sparger air rates and pressures,
- characterize oxidation system performance versus plant operating parameters (sulfur level, pH, etc.),
- develop complete guidelines for dewatering equipment (thickeners, centrifuges, and rotary drum vacuum filters) in the forced oxidation operating mode.

In March 1983 three towers on Unit 1 were placed into service in the forced oxidized mode. Research-Cottrell operating and supervisory personnel, with the assistance of plant operations and laboratory personnel, began an around-the-clock monitoring and testing program. Oxidation equipment performance was monitored and a "fine tuning" program was begun to optimize oxidation air pressures and rates.

By mid April the go-ahead was given to put Unit 2 on line with six operating towers equipped with spargers. Oxidation levels, as measured in the thickener underflow, steadily increased beyond the 90% levels. Test activities were initiated on the vacuum filters and centrifuges. A characterization program to evaluate the effect of process variables on oxidation efficiency versus operating pH, inlet SO_2 levels, and oxidation air rates was undertaken. Installation of spargers into the remaining towers was completed by mid June, and these towers were put on line.

The key findings of the characterization program were:

- The extensive collection of slurry samples for analysis identified the importance of proper operation (and maintenance) of FGD towers. Excessive carbonate levels in the thickener were found to be a result of operation of partially plugged towers, and overflowing absorber feed tanks.

- Consistently high levels of oxidation could be maintained at the tested sulfur levels.

- The vacuum filters were capable of exceeding design levels of filter cake production.

Additional observations are provided below in the discussion of plant operating experience.

OPERATING EXPERIENCE AT MARTIN LAKE TO DATE

Operating experience of the various pieces of equipment that are affected by (or affect) the production of a high quality gypsum by-product at Martin Lake is discussed below.

Scrubber Tower Performance

The need to stabilize the chemistry in the scrubber towers led to further process modifications, instrumentation upgrades, and a more comprehensive chemical quality control program. Texas Utilities, Research-Cottrell, EPRI, and various consultants worked together in a task force to design, test and incorporate these changes. This program is the subject of other reports; only portions that affect gypsum by-product quality will be mentioned.

Quencher/Oxidation Sump pH Control--

The pH of the quencher/oxidation sump is known to be a critical factor in achieving maximum oxidation. The results of the 1983 testing defined a need to maintain a sump pH of 5.0 to 5.3 to optimize oxidation, quencher section SO_2 removal and residual $CaCO_3$ levels. Texas Utilities implemented a control system for each tower quencher sump based on the task force recommendations. This additional pH control loop supplements the absorber loop pH controls originally supplied.

pH Monitoring Instrument Upgrades--

In order for the pH control of the quencher sumps to be effective, higher reliability of the sump pH monitoring instruments was necessary. The existing TBI 600 Series pH system, which has been in service since 1979 on all towers, was prone to problems with electrical connectors, bulb electrode malfunction, and signal noise between local preamp and remote processor. At the recommendation of the task force, Texas Utilities purchased TBI 513 Series pH systems. This series locates the complete electronic package at the sampling point and transmits a 4 to 20 m.a. signal to the control room. To date, this equipment has performed well.

Since the startup of these new pH control loops and pH instruments in the fall of 1984, the chemistry in the scrubber tower has been tightly controlled. Oxidation of sulfite to sulfate has consistently been above 95%. Reagent utilization has been consistently above 90% and tower SO_2 removal has remained consistently above 90%.

Oxidation Air Sparger

One of the key components in the system design is the oxidation air sparger. The function of this device is to ensure an intimate contact between oxygen (from the air) and the sulfite in solution. A periodic manual water flush system was incorporated to permit on-line cleaning of the sparger piping inside each tower.

Mechanical Failures--

Three to six months after the spargers were placed in service weld failures were observed on the spargers in three towers. Upon investigation, it was determined that the U-bolts used to secure the sparger piping to the tower sump floor brackets had failed due to vibration and corrosion. The sparger laterals, once free of the bracket, apparently generated ample vibration to cause the weld between the lateral branches and the main header to fail. This problem was solved by wrapping the pipe with 1/8" thick rubber at each U-Bolt to dampen sparger vibration transmitted to the U-Bolt threads and to double-nut the U-Bolts using nylon loaded locking nuts.

Pluggage--

Soon after startup several incidences of sparger plugging were experienced. Frequent on-line flushing was not effective in controlling plugging, and in some cases appeared to make things worse. The cause of the plugging was identified as three separate problems: (1) plugging of the holes by solids, (2) scale formation, and (3) burying spargers with slurry solids.

1. The water source utilized for manually flushing the spargers was found to contain large amounts of solid debris. Periodic flushing of the headers introduced these solids into the sparger pipe causing some holes to plug. Upgrading the quality of this wash water has eliminated this problem.

2. Soon after startup, a gradual buildup in sparger pressure drop was observed in several towers. Manual water flushing was not effective in eliminating this buildup. The problem was identified as scale growth over the holes inside the pipe. The exact mechanism was not determined but it appeared

to be dependent upon flush water quality, temperature of the gas-liquid-metal interface and the "arching diameter" of the gypsum crystals. This problem was eliminated by enlarging some holes. To date, this modification has eliminated this problem.

3. The problem of sparger headers being buried by solids occurred on several occasions after startup. This problem was a result of insufficient agitation in the quencher sump and occurred in several of the original 18 towers. The six new towers, which were supplied with an upgraded 3-agitator design, did not have this problem. The original eighteen towers are presently being upgraded to incorporate the new agitator design.

The modifications mentioned above, along with manual air flow regulation and manual flushing of the header every six hours, has provided satisfactory sparger operation over the past year.

Oxidation Air Compressors

One of the major costs of the Martin Lake forced oxidation upgrade, in terms of both capital and operating expense, was the oxidation air compressors. Martin Lake was provided with four Seimens Allis, single stage centrifugal air compressors. Each compressor is driven with a 1750 HP motor and provides up to 45,700 ACFM of air.

During startup activities, problems with compressor controls and instrumentation were identified and resolved by Texas Utilities and Research-Cottrell. The experience showed the complexity of operating up to four compressors, feeding a common header which can feed different combinations of up to 24 towers. With all three units in operation, each with five scrubber towers on line, three operating compressors are required. Each compressor operates at about 65% of rated airflow and motor horsepower capacity.

To date, these compressors have demonstrated reliable performance. However, due to the high energy consumption of these compressors, Texas Utilities is planning to optimize compressor operation to provide air requirements economically to the sparger.

Thickeners

The advent of forced oxidation and its inherent changes in the characteristics of the scrubber solids has affected the thickeners more than any other piece of equipment in the FGD system. These effects have caused changes to the operating and maintenance procedures involving the thickeners and their support equipment.

As expected, forced oxidation dramatically improved the efficiency of the thickeners as a dewatering device. Prior to forced oxidation, the thickener overflow typically contained 5% to 7% solids. This condition necessitated the use of a coagulating polymer to bring the solids down to the 2% to 4% level, which could be tolerated by the scrubbers as process dilution water. The same thickeners now operate at less than 1% solids in their overflow (with no polymer) due to the improved settling tendency of oxidized FGD sludge.

The negative aspect of thickener operation with oxidized sludge is that a given thickener is much less "forgiving" to solids overloading than it was with unoxidized sludge. The forced oxidation upgrade provided increased thickener underflow velocity to compensate for the increase in underflow solids. Thickener underflow solids increased from about 30% (for unoxidized sludge) to 35% to 45% with oxidation. Dilution water and an air sparger were provided in the thickener to maintain a constant feed density to the secondary dewatering equipment. These additional features do their job well; however, problems can still develop. Loss of the underflow pumps for longer than 30 minutes will normally result in the plugging of underflow. The loss of the thickener rake for more than 30 minutes can also result in problems. After the rake has been stationary for a short period of time, the solids will actually bury the rake plows and superstructure. The rake drive then cannot overcome the additional loading if started again. The thickener must then be drained and the solids washed off the rake with high pressure hoses.

It has been found that a small amount of added attention and quick reaction can keep these problems from occurring. The great improvement in thickener performance which forced oxidation has provided, is well worth the added attention.

Vacuum Filters

The utilization of the gypsum produced by the forced oxidation upgrade will depend heavily on the successful operation of the four Komline-Sanderson, 10 foot diameter, rubber-lined, rotary drum vacuum filters. These filters have demonstrated dependable operation since the completion of revisions made during the startup period. Problems encountered and resolved during startup centered in two main areas: (a) mechanical drives and (b) filter cloth failures.

Mechanical Drives--

> The drive problems involved the vat agitator drive gearbox and motor, and the vacuum drum drive. Upon initial operation, it became apparent that the settling characteristics of the gypsum were going to be better than originally expected. This fact, coupled with continuous service requirements of the vat agitator, led to a continuing stallout of the vat agitator drive. Increasing drive motor horsepower improved the situation, but resulted in periodic shearing of the gear drive keys. Resolution of this problem came with the upgrading of both the drive gearbox and the motor.

The variable speed D.C. drive from the vacuum drums also suffered from stall problems and component failure. It was determined that the drive motors were overheating due to the drum filter being able to pick up more pounds of gypsum in a revolution than was originally expected. This problem was resolved with the installation of cooling fans on the drive motors and the identification of minimum operating speeds for the drum drives.

Filter Cloth Failures--

It has been determined at Martin Lake that by far the most critical component in the dewatering of gypsum is the filter cloth. Numerous instances of abrasion of drum filter internals, abrasion of filtrate valve assemblies and high filtrate liquid levels in the vacuum pump can all be traced directly to holes in the filter cloth. Moisture content of the gypsum cake is also affected by holes in the cloth.

Investigations into more durable cloths have identified nylon as a very durable material. Testing will continue at Martin Lake to optimize durability and filtering performance of filter cloth.

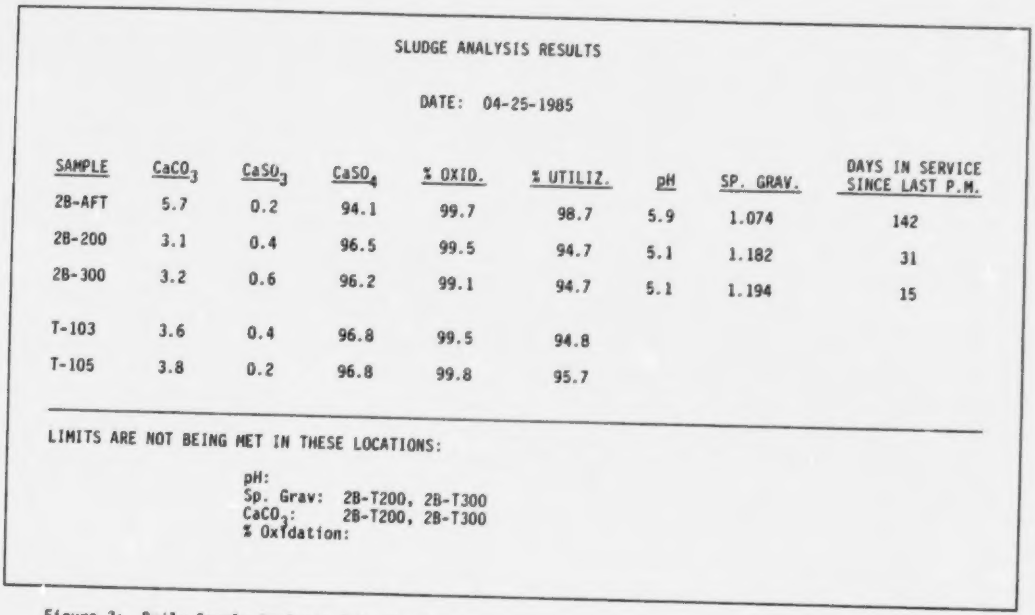
From an operational standpoint, the vacuum filters appear to be very reliable. Expedient cloth repair and/or replacement seems to be the key to reliability. As with all equipment handling slurries, piping design of drain and flush systems is important since the potential of plugging always exists.

In recent months, filter cloth blinding has been noted. The blinding appears to be caused by very small gypsum crystals. The source of these small crystals is presently being investigated. Chemical problems in the quencher/oxidation sump and the concentrating of small crystals by the existing dewatering centrifuges may be the causes of these blinding problems.

ANALYTICAL PROGRAM AND PERFORMANCE DATA

Texas Utilities developed a comprehensive analytical program to support plant operations and to monitor gypsum by-product quality. This program is structured to provide rapid feedback of analytical results. This allows operating personnel to take corrective actions as necessary. The goals of this program are to maintain operating conditions to maximize SO_2 removal, reagent utilization and to produce a high quality, saleable by-product.

At the present time the solid phase chemistry of the FGD system is monitored. Samples of slurry are collected daily from each operating quencher sump, absorber feed tank and thickener. The pH and specific gravity of each sample are measured and recorded. The solids are then analyzed to determine their composition. A computer printout of the daily analysis is issued about 4 hours later. Samples which do not meet certain composition or operating limits are identified. Figure 3 shows a portion of a typical daily analysis sheet. In the near future this analytical program will be expanded to include liquid phase analysis.



SLUDGE ANALYSIS RESULTS

DATE: 04-25-1985

SAMPLE	$CaCO_3$	$CaSO_3$	$CaSO_4$	% OXID.	% UTILIZ.	pH	SP. GRAV.	DAYS IN SERVICE SINCE LAST P.M.
2B-AFT	5.7	0.2	94.1	99.7	98.7	5.9	1.074	142
2B-200	3.1	0.4	96.5	99.5	94.7	5.1	1.182	31
2B-300	3.2	0.6	96.2	99.1	94.7	5.1	1.194	15
T-103	3.6	0.4	96.8	99.5	94.8			
T-105	3.8	0.2	96.8	99.8	95.7			

LIMITS ARE NOT BEING MET IN THESE LOCATIONS:

pH:
Sp. Grav: 2B-T200, 2B-T300
$CaCO_3$: 2B-T200, 2B-T300
% Oxidation:

Figure 3: Daily Sample Analysis Data Sheet

A summary of the analytical data obtained from December 1984 to May 1985 is shown in Figures 4 and 5. This data shows the daily chemical analysis of the solids in the thickener underflow during this period. It is important to note that slurry fed to the thickener consists of all scrubber waste products. In addition to oxidation sump slurry, the feed stream to the thickener contains slurry from infrequent process upsets (such as an overflowing absorber feed tank) and waste slurry from regular tower maintenance. However, since the thickeners provide a large surge volume these additional discharge streams are not expected to have a detrimental affect on the composition of the gypsum. In the case of a unit outage, which would result in an out of spec. gypsum, the underflow from the thickener would be diverted to disposal.

Figure 4 shows the concentration of unreacted calcium carbonate in the thickener underflow. The average concentration of calcium carbonate in the solids has been about 6% during this period. The data points showing calcium carbonate levels above 14% occured during startup of the FGD system.

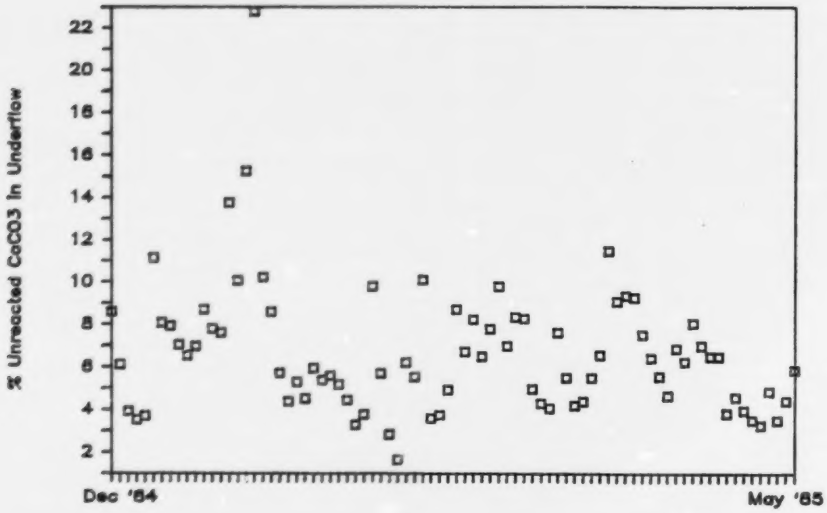


Figure 4 - % Unreacted $CaCO_3$ in thickener underflow

Figure 5 shows the concentration of calcium sulfate (gypsum) in the thickener underflow for this same time period. Typically the underflow solids contain greater than 90% calcium sulfate and more recently have averaged closer to 95%. The data showing very low oxidation again corresponds to startup of the FGD system.

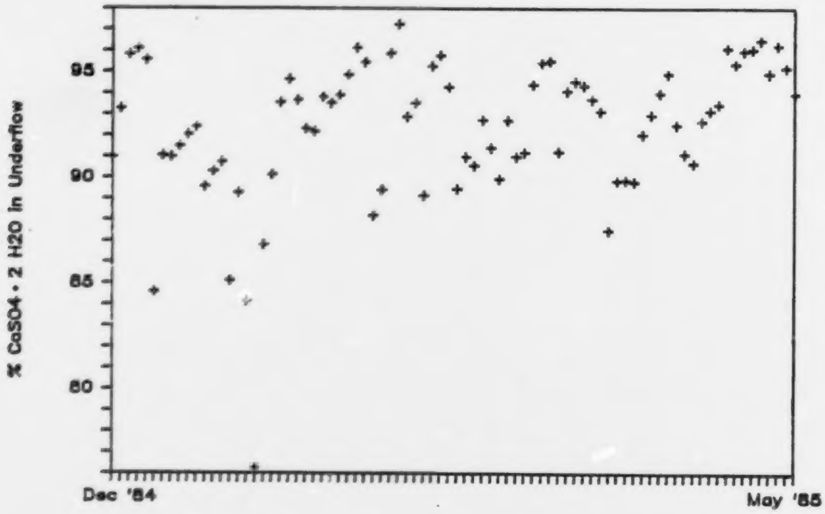


Figure 5 - % $CaSO_4 \cdot 2H_2O$ in thickener underflow

The concentration of chlorides in the FGD system is monitored regularly. Besides the adverse effects which chlorides are known to have on FGD, the presence of high concentration of chlorides will impact the potential for gypsum sales. Operation of the wastewater treatment equipment at Martin Lake has maintained low chloride concentrations. Typically the thickener liquid phase chloride concentration is 250 ppm or less.

THE BY-PRODUCT GYPSUM SALES AGREEMENT AND WALLBOARD MANUFACTURING PLANT

As the startup of the forced oxidation upgrade was being completed, it became obvious that a potential to produce saleable gypsum existed. Texas Utilities began to explore markets for gypsum in much the same way that fly ash and bottom ash markets were explored and found previously.

On January 16, 1984, Texas Utilities entered into an agreement with Windsor Properties, Inc. for testing and development to determine the potential for Martin Lake gypsum to be used in the manufacture of wallboard. A subsidiary of Windsor Properties, Windsor Gypsum, Inc., was actually involved in the testing and development. The agreement provided for a test period in which Texas Utilities agreed to supply Windsor Gypsum with up to 1,000 tons of by-product gypsum. Windsor, in turn, agreed to perform research and development, make production runs of wallboard from this gypsum, and perform marketing surveys. During the test period, several hundred square feet of wallboard was produced by Windsor Gypsum. Table 3 shows the chemical analysis of this gypsum.

TABLE 3. CHEMICAL ANALYSIS OF MARTIN LAKE SYNTHETIC GYPSUM

Moisture, %	13.8 - 17.5
Purity, %	88.3 - 93.3
MgO, %	0.06 - 0.07
Na_2O, %	0.018 - 0.046
Cl, ppm	28 - 47
SO_2, %	0.010 - 0.016
K_2O, %	<0.013
P_2O_5, %	0.050 - 0.064
F , %	0.002 - 0.003
pH	6.1 - 6.2
Bulk Density, kg/m^3	1,105 - 1,120
$CaCO_3$, %	4.8 - 12.5

The range in listed values is based on analysis of 4 samples.

As a result of this testing and various other tests by consultants employed by both Texas Utilities and Windsor Gypsum, Inc., and by Texas A and M University, a Synthetic Gypsum Sales Agreement between Texas Utilities and Windsor Properties, Inc. was executed on October 16, 1984. The agreement calls for the delivery of approximately 30,000 tons per month of Martin Lake by-product gypsum to a wallboard plant to be located adjacent to a Texas Utilities owned railroad line. The gypsum will be delivered by the same Texas Utilities railcars that had previously been hauling the material to the permitted disposal areas.

On December 6, 1984, the following press release was made by Windsor Gypsum, Inc., Henderson Division:

> J. P. Boone, Chairman and Chief Executive Officer of Windsor Gypsum, Inc., announced today that a new wallboard manufacturing plant will be built in Rusk County, eight miles northeast of Henderson, just south of State Highway 43. Boone said the facility will use the most ultra-modern state-of-the-art equipment available, and will be the first plant in the United States to produce wallboard from manufactured gypsum.
>
> The facility, to be erected on a 67-acre tract of land, will include a 200,000 square-foot fireproof steel building and a 4,000 square-foot colonial style brick office building.
>
> Boone said, "When we are in full production, we will receive 1,000 tons of gypsum per day. The plant is designed to produce 4' x 8' x ½" wallboard at a rate of 250 feet-per-minute, or 1,500,000 square feet in a 24 hour period. This is the equivalent of 60 truckloads of finished wallboard ready for the market per day".
>
> "Manufactured gypsum dries faster and is stronger than natural gypsum," Boone added. He stated that his tests have shown that wallboard made from gypsum from Martin Lake is stronger than wallboard made from natural gypsum.

Windsor will purchase gypsum produced by Texas Utilities Generating Company's Martin Lake generating plant near Tatum.

CONCLUSION

Forced oxidation was designed into the Martin Lake FGD system primarily to reduce sludge handling equipment requirements and provide the most cost effective design for the system upgrade. The potential for by-product sales was identified early in the design, and provisions were made to allow for this potential.

Since Texas Utilities already sold some by-products from lignite combustion a market was explored for the FGD by-product. The potential for savings in disposal costs as well as the generation of income to offset operating costs was determined to be significant, thus providing the impetus for finding a gypsum market.

In the final analysis the operation of the Martin Lake FGD system as "gypsum production units" is compatable with operation of the system as flue gas scrubbers. The activity required to maintain good quality gypsum has been found to improve the operation and enhances the performance of the FGD system.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore, the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

EVALUATION OF ENGINEERING PROPERTIES AND WET STACKING DISPOSAL OF WIDOWS CREEK FGD GYPSUM-FLY ASH WASTE

by

John E. Garlanger, Ardaman & Associates, Inc., Orlando, Florida
Sal H. Magliente, Tennessee Valley Authority, Chattanooga, Tennessee
Thomas S. Ingra, Ardaman & Associates, Inc., Orlando, Florida
James L. Crowe, Tennessee Valley Authority, Chattanooga, Tennessee

ABSTRACT

Wet stacking of by-product gypsum has been practiced by the phosphate fertilizer industry for more than 25 years. The ability to use wet stacking for disposal of flue gas desulfurization (FGD) gypsum was first demonstrated during an Electric Power Research Institute sponsored project on Chiyoda Thoroughbred 121 FGD gypsum produced at the Scholz Electric Generating Station of Gulf Power Company in Sneads, Florida. Wet stacking of FGD gypsum containing fly ash, however, has not been previously demonstrated. Accordingly, as part of an overall project investigating various FGD waste disposal alternatives, the Tennessee Valley Authority constructed a pilot-scale wet stacking disposal facility to evaluate the feasibility of wet stacking FGD gypsum-fly ash waste produced at the Widows Creek Steam Plant in Stevenson, Alabama. Operational experience and results from geotechnical laboratory testing performed on the waste are presented. The results indicate that although the Widows Creek FGD gypsum-fly ash had settling, dewatering, and structural characteristics not as favorable for stacking as phosphogypsum or CT 121 FGD gypsum, they were adequate for wet stacking. Therefore, the project findings should extend the ability of the utility industry to employ wet stacking disposal to facilities which also use FGD/forced oxidation systems as the primary particulate removal process.

INTRODUCTION AND OVERVIEW OF WIDOWS CREEK STEAM PLANT

The Tennessee Valley Authority (TVA) has been developing scrubber operating and waste disposal experience with limestone flue gas desulfurization (FGD) systems via a series of demonstration projects on Unit 8 of the Widows Creek Steam Plant in Stevenson, Alabama (1) (2) (3). Unit 8 is a coal-fired 550-MW boiler which became operational in 1964, and is the newest of eight coal-fired units at Widows Creek. With its sister unit, Unit 7, it occupies a separate powerhouse about one-quarter mile from the powerhouse containing Units 1 through 6. The aggregate capacity of the eight units is 1,950 MW. A schematic site plan of the Widows Creek Steam Plant is shown in Figure 1.

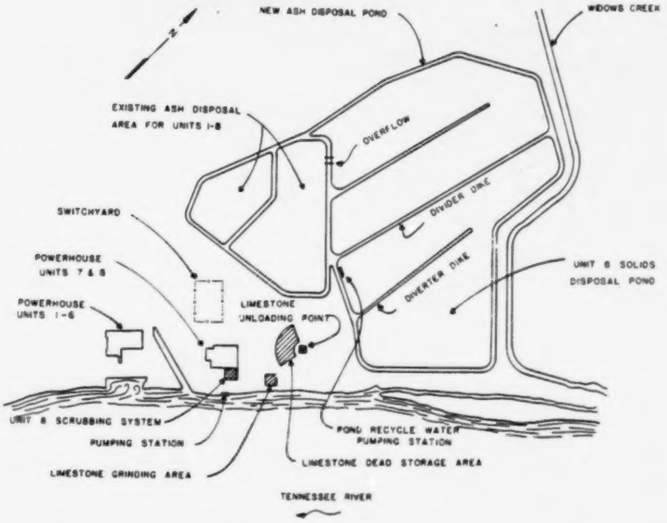

N
NEW ASH DISPOSAL POND
WIDOWS CREEK
EXISTING ASH DISPOSAL AREA FOR UNITS 1-8
OVERFLOW
DIVIDER DIKE
SWITCHYARD
POWERHOUSE UNITS 7 & 8
DIVERTER DIKE
UNIT 8 SOLIDS DISPOSAL POND
POWERHOUSE UNITS 1-6
LIMESTONE UNLOADING POINT
UNIT 8 SCRUBBING SYSTEM
PUMPING STATION
POND RECYCLE WATER PUMPING STATION
LIMESTONE DEAD STORAGE AREA
LIMESTONE GRINDING AREA
TENNESSEE RIVER

Figure 1. Widows Creek Steam Plant Site Plan.

Various landfilling methods involving dewatering, stabilization, and fixation of FGD calcium sulfite wastes produced by Unit 8 have been studied by TVA. Primary among these studies was the use of dry fly ash for stabilization of dewatered FGD calcium sulfite waste and landfill disposal. This methodology did not appear feasible for Unit 8, because insufficient dry fly ash was available for stabilization.

Subsequently, one of the four modules of the Unit 8 FGD system was modified to permit forced oxidation of calcium sulfite to calcium sulfate (gypsum), which is a waste considered more suitable for disposal. As part of an evaluation of various disposal alternatives for FGD gypsum-fly ash waste, a wet stacking disposal demonstration project was initiated. Gypsum produced as a by-product at phosphate fertilizer plants, i.e., phosphogypsum, is disposed of using the wet stacking method, and hence it was anticipated that FGD gypsum-fly ash waste could also be disposed of using the wet stacking method.

The objectives of the demonstration project were to evaluate the engineering properties of FGD gypsum-fly ash waste produced by the scrubber relevant to wet stacking, and to evaluate the feasibility of using the wet stacking method for a full-scale disposal facility at Widows Creek. Specifically, the objectives of the project were to:

- Demonstrate the construction and operation of a pilot-scale wet stacking disposal facility for FGD gypsum-fly ash waste.
- Determine the handling, stackability and trafficability characteristics of the gypsum-fly ash waste and identify potential wet stacking disposal construction and operation problems.
- Develop a supporting data base of engineering properties from laboratory and field testing for use in the design of a full-scale FGD gypsum-fly ash wet stacking disposal facility.

This paper presents the engineering properties of Widows Creek FGD gypsum-fly ash waste relevant to wet stacking and field observations obtained during construction and operation of a 0.3-acre, 12-foot high pilot scale wet stacking disposal facility at the Widows Creek Steam Plant. The findings of this study were used as the basis for design of a 120-acre wet stacking disposal facility for Units 7 and 8 which is scheduled to become operational in the fall of 1985.

DESCRIPTION OF WIDOWS CREEK FORCED OXIDATION FGD LIMESTONE SCRUBBER SYSTEM

A FGD limestone scrubber system was installed on Unit 8 of the Widows Creek Steam Plant. Unit 8 is a 550-MW coal-fired boiler employing four identical parallel FGD trains, namely trains A, B, C and D, that scrub flue gas from the

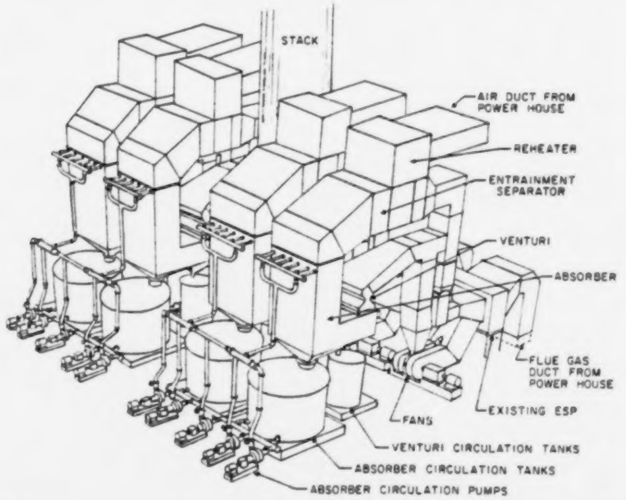

STACK
AIR DUCT FROM POWER HOUSE
REHEATER
ENTRAINMENT SEPARATOR
VENTURI
ABSORBER
FLUE GAS DUCT FROM POWER HOUSE
EXISTING ESP
FANS
VENTURI CIRCULATION TANKS
ABSORBER CIRCULATION TANKS
ABSORBER CIRCULATION PUMPS

Figure 2. Widows Creek Steam Plant Unit 8 Scrubber Facility.

boiler. One FGD train is connected to each of the four flue gas ducts from the boiler. A schematic illustration of the Unit 8 FGD scrubber system is presented in Figure 2.

As shown, the FGD scrubber consists of an adjustable throat venturi followed by a grid-type absorber spray tower. For the forced-oxidation demonstration project, the D train was modified to allow forced oxidation of the scrubber sludge by installing air-sparging and turbine agitation equipment in the D train absorber and venturi circulation tanks.

The forced oxidation of calcium sulfite to calcium sulfate requires: aeration of the scrubber slurry in the absorber and venturi circulation tanks at atmospheric pressure; and maintaining a suitable pH in the slurry at the point of air introduction. Since the scrubber was also used as the primary particulate removal system, the resulting waste was a combination of calcium sulfate and fly ash. Unreacted limestone and unoxidized calcium sulfite also occurred within the waste.

Aeration in the absorber and venturi circulation tanks was provided by four screw type diesel air compressors. The compressors were rated at 1,600 acfm and 100 psig at the outlet. Air flow to the absorber and venturi circulation tanks was controlled by a rubber pinch valve at the inlet to the tanks. The venturi circulation tank was 18 feet in diameter and 25 feet high with a working capacity of 40,700 gallons (Figure 3). The absorber circulation tank was 33 feet in diameter and 25 feet high with a working capacity of 147,000 gallons (Figure 3).

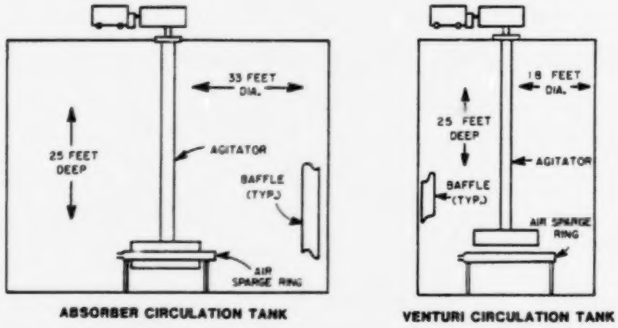


Figure 3. Absorber and Venturi Circulation Tanks.

Normally, only two compressors were needed for complete oxidation, with the other two compressors serving as standby units. The compressed air was distributed within the absorber and venturi circulation tanks by means of a sparge ring (Figure 4) consisting of a perforated annular duct surrounding the agitator impeller. The sparge rings were held above the tank floor by supports welded to the floor.

The turbine agitators installed inside and directly above the sparge rings in the absorber and venturi tank, respectively, were significantly larger than standard circulation tank agitators to enhance mixing. The bladed turbine agitators were constructed of carbon steel and coated with rubber. The agitators generated turbulent eddies which sheared the air bubbles emitted from the sparge ring. This provided a high gas/liquid mass transfer area, increased oxygen uptake by the slurry, and prevented gypsum crystals from settling.

Limestone was added to adjust the pH as necessary to maintain a suitable pH for SO_2 removal and yet achieve complete oxidation. Limestone was ground by a wet ball mill before preparation into a slurry. The limestone slurry was stored in one tank feeding the four FGD trains. The limestone slurry tank was 36 feet in diameter and 27 feet high with a working capacity of 193,000 gallons. The limestone slurry tank was equipped with an agitator to mix the slurry and to keep limestone particles in suspension. The limestone slurry was pumped to the absorber circulation tank by slurry feed pumps, one pump for each of the four FGD trains.

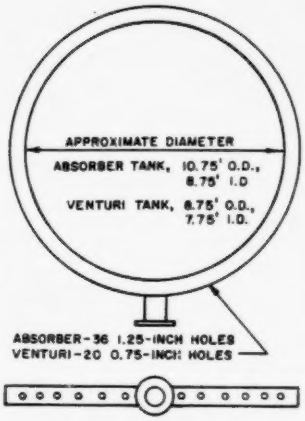

APPROXIMATE DIAMETER
ABSORBER TANK, 10.75' O.D., 8.75' I.D
VENTURI TANK, 8.75' O.D., 7.75' I.D.
ABSORBER-36 1.25-INCH HOLES
VENTURI-20 0.75-INCH HOLES

Figure 4. Air Sparge Ring.

The pH in the absorber circulation tank was regularly monitored. The slurry pH was adjusted by varying the limestone slurry flow into the absorber circulation tank. Under normal conditions, the optimum pH in the absorber circulation tank for complete oxidation was 5.8.

Oxidized FGD gypsum-fly ash waste from the scrubber was transported as a slurry to the wet stack disposal area by piping connected to an 8-inch gate valve near the bottom of the venturi circulation tank. A 250-gallon per minute pump was initially installed to pump the slurry from the venturi circulation tank to the disposal area, which was located approximately one-quarter mile from the scrubber facility. However, during the majority of the test program the slurry flowed by gravity to the stacking area because the pump drew the venturi circulation tank level down too fast, resulting in the loss of the venturi slurry seal which allowed partially scrubbed flue gas to escape. Eventually, the pump was abandoned and an automatic control system was installed to monitor the venturi circulation tank slurry level. Gravity flow was then stopped automatically when the slurry level dropped to a predetermined level.

After clarification in the stacking area, the process water was decanted into a sump area and pumped to the existing Unit 8 scrubber pond (Figure 1). The pump operated on a float switch which controlled the water level in the sump.

ENGINEERING PROPERTIES OF WIDOWS CREEK FGD GYPUM-FLY ASH WASTE

As part of the evaluation of the feasibility of wet stacking FGD gypsum-fly ash waste produced at Widows Creek, a laboratory testing program was performed to characterize the engineering properties of the waste relevant to wet stacking disposal, and to provide properties for use in the design of a full-scale facility (4). The chemical composition, crystal morphology, and physical characteristics of the waste are first addressed. Specific engineering properties regarding settling, consolidation, permeability and shear strength are subsequently discussed.

CHEMICAL COMPOSITION AND CRYSTAL MORPHOLOGY

The chemical compositon of a FGD gypsum-fly ash waste is expected to influence its engineering properties. Generally, the presence of calcium sulfite, due to reduced oxidation levels, inhibits dewatering, decreases settling rates, increases compressibility, and decreases permeability and shear strength. Further, high fly ash content wastes are expected to be less suitable for wet stacking than those higher in gypsum content.

During operation of the scrubber facility, the chemical composition of gypsum-fly ash solids within the venturi bleed stream was typically determined three times daily. The chemical composition of the venturi bleed stream solids was characterized by the following average constituents:

- 41% gypsum
- 43% fly ash
- 13% limestone
- 0.7% sulfite

As shown, the chemical composition was characterized by essentially equal weights of gypsum and fly ash with 13% unreacted limestone and 0.7% unoxidized calcium sulfite. The low calcium sulfite content resulted from the high average degree of oxidation achieved of 97.9%. The fly ash content never decreased below 26.6%, and occasionally reached as high as 75.6%. The gypsum content never exceeded 61.6%, and decreased to as low as 5.9%. Crystalline phases detected from X-ray diffraction analyses were gypsum, magnetite, hematite, calcite, dolomite and quartz, which are consistent with the constituents found via wet chemistry.

TVA anticipates that FGD gypsum-fly ash waste generated from the full-scale forced oxidation scrubber on Units 7 and 8 will have an average chemical composition of: 60% gypsum, less than 30% fly ash, less than 10% unreacted limestone and less than 1% unoxidized calcium sulfite. Accordingly, the full-scale facility is projected to produce waste higher in gypsum and lower in fly ash than produced during the demonstration project, since fly ash will be removed in Unit 7 prior to scrubbing.

The crystal morphology of the Widows Creek FGD gypsum-fly ash waste was determined via scanning electron microscopy. The typical crystal morphology found for the gypsum-fly ash waste is illustrated in Figure 5.

As shown, the major features visible in the photomicrograph include lath-shaped gypsum crystals, spheres (fly ash) with a wide range of sizes, and a general "dusty" appearance on the gypsum crystals and fly ash spheres.

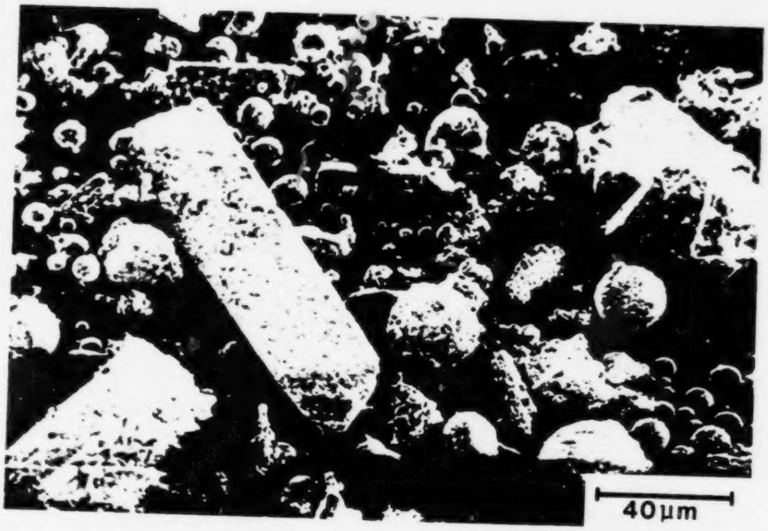


Figure 5. Scanning Electron Photomicrograph of FGD Gypsum-Fly Ash Waste.

PHYSICAL CHARACTERISTICS

The Widows Creek FGD gypsum-fly ash waste was found via sieve and hydrometer analyses to be predominantly a silt-sized material. As shown by the particle size distribution in Figure 6, the gypsum-fly ash waste exhibited a range of 64 to 94% and an average of 76% by dry weight of the particles finer than the 74 µm size. Less than 10% of the particles by dry weight were finer than the 45 µm clay particle size. Most of the plus 74 µm size particles are unreacted

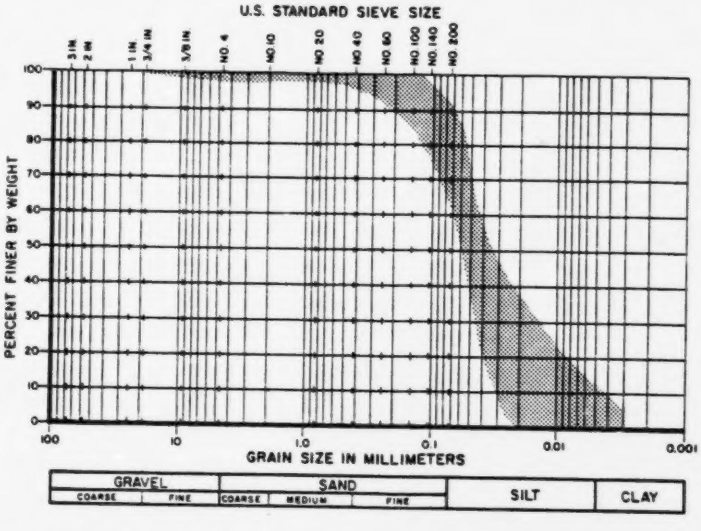


Figure 6. Particle Size Distribution of FGD Gypsum-Fly Ash Waste.

limestone, with the gypsum, fly ash and unoxidized calcium sulfite phases occurring within the minus 74 μm size fraction.

The particle size distribution is similar to that previously reported for Widows Creek gypsum-fly ash waste (5), and is also consistent with particle size distributions typically reported for FGD gypsum (6)(7) and fly ash (6).

The specific gravity of the gypsum-fly ash waste was determined to vary from 2.34 to 2.93 with an average of 2.50±0.14. This average value is in general agreement with the specific gravity of 2.46 reported for a waste with 25% gypsum and 75% fly ash (7) and 2.29 reported for a waste with 55% gypsum, 20% limestone, 25% fly ash, and 1% calcium sulfite (5). For comparison, the specific gravity of the various minerals within the waste are: gypsum (2.33); calcium sulfite (2.50); magnetite (5.18); hematite (5.25); calcium carbonate (2.70); magnesium carbonate (2.85); and quartz (2.66). A much higher specific gravity is calculated using the mineral specific gravities and the weight percent of each mineral within the waste. The measured average gypsum-fly ash waste specific gravity of 2.50, however, is consistent with the expected value using a fly ash specific gravity of 2.50 as often reported. The fly ash specific gravity is typically lower than its constituent mineral specific gravities, due to the "porous" nature of the spheres.

SEDIMENTATION CHARACTERISTICS

Settling tests performed on the venturi bleed stream solids indicated that the settling rate, Q, varied from a minimum of 0.32 cm/min, for a poor level of oxidation (62.9%), to a maximum of 4.8 cm/min with an overall average of 3.0 cm/min. The effect of oxidation of solids on the settling rate is presented in Figure 7. As expected, although the data are limited, the settling rate decreases markedly below an oxidation level of approximately 90% when the calcium sulfite content exceeds about 3% (Figure 8).

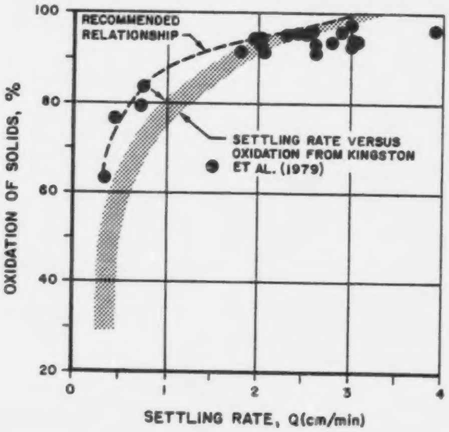


Figure 7. Settling Rate Versus Oxidation of Solids.

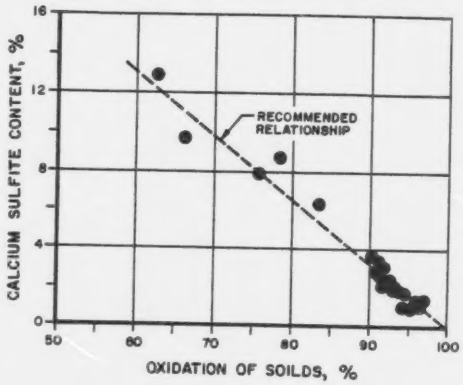


Figure 8. Oxidation of Solids Versus Calcium Sulfite Content

The settling rate versus oxidation of solids previously found for Widows Creek FGD gypsum-fly ash waste (8) is also included on Figure 7 for comparison. As shown, the correlations are similar.

The "final" settled solids content, S_F, determined at the end of the settling tests when no additional change in height was observed ranged from 44.6% to 84.1%, with an average of 63.5%. Accordingly, an average solids content of 63.7% is initially expected in the disposal area after sedimentation, but before consolidation under subsequent layers of sedimented waste. The average "final" settled solids content corresponds to a moisture content of about 57%. For a specific gravity of 2.50, the "final" settled solids content corresponds to a dry density of 64.3 lb/ft^3 and a total saturated unit weight of 101.0 lb/ft^3.

The presence of calcium sulfite in FGD wastes, due to poor solids oxidation, decreases the settling rate and "final" settled solids content. These effects are illustrated in Figure 9. Although the data are limited at higher calcium sulfite contents, the trends of decreasing settling rate and decreasing "final" settled solids content with incresing calcium sulfite content are clear. The trends also indicate that relatively small percentages of unoxidized calcium sulfite can affect settling characteristics. Gypsum-fly ash waste with 3% calcium sulfite, which corresponds to about 90% oxidation (Figure 8), results in a reduced settling rate of

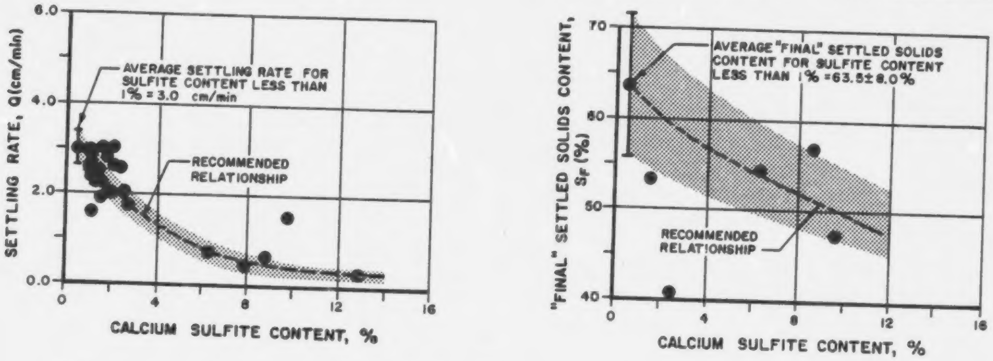


Figure 9. Effect of Calcium Sulfite Content on Settling Rate and "Final" Settled Solids Content.

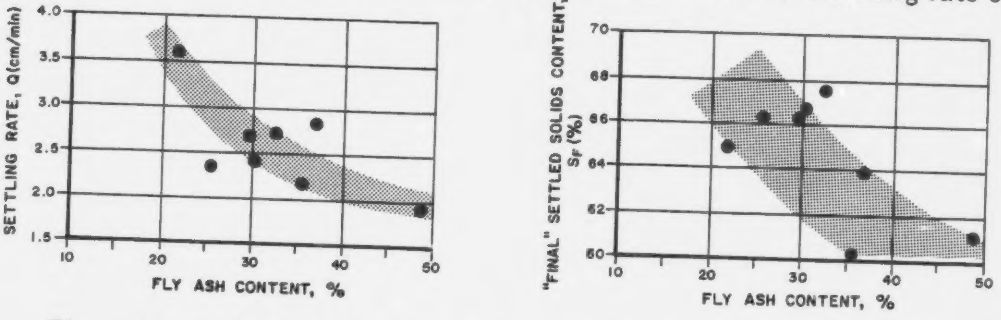


Figure 10. Effect of Fly Ash Content on Settling Rate and "Final" Settled Solids Content.

approximately 1.5 cm/min rather than 3.0 cm/min, and a "final" settled solids content of 58% rather than 63.5%.

Correlations between gypsum and fly ash content with settling rate and "final" settled solids content were attempted. No definitive correlations were apparent. As shown in Figure 10, however, weak trends of decreasing settling rate and decreasing "final" settled solids content with increasing fly ash content were observed.

COMPRESSIBILITY AND CONSOLIDATION CHARACTERISTICS

The void ratio, e, (or solids contents) versus effective vertical consolidation stress, $\bar{\sigma}_{vc}$, curve developed for the FGD gypsum-fly ash waste from the slurry-consolidation and settling tests is presented in Figure 11. As shown, the void ratio, e, ranges from 1.3 (S=65.9%) at low effective stresses typical of self-weight consolidation stresses existing after sedimentation, to 0.80 (S=75.8%) at an effective stress of 4.0 kg/cm^2 which corresponds approximately to the effective stress existing at the base of a 150-foot high stack. The compression ratio, or slope of the void ratio versus effective stress curve expressed in terms of strain is approximately 0.05 between stresses of 0.01 to 1.0 kg/cm^2. Comparatively, this compression ratio is similar to that expected for a loose sand. At low stresses between 0.001 to 0.01 kg/cm^2, i.e., below less than 1 foot of material, the material is relatively more compressible, with a compression ratio of 0.14.

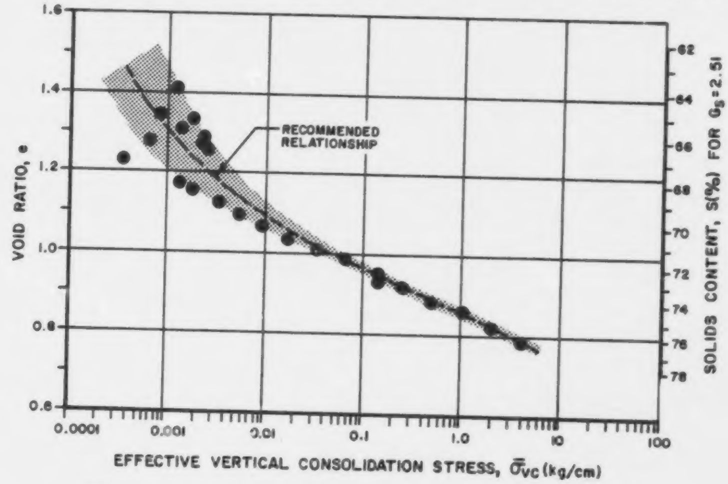


Figure 11. Void Ratio Versus Effective Stress Compressibility Curve.

Typically, FGD gypsums consolidate quickly under self-imposed or externally applied loads. Following consolidation, drained creep or secondary compression occurs resulting in slight increases in dry density with time at a given effective stress. The Widows Creek FGD gypsum-fly ash waste was also found to consolidate quickly with representative coefficients of consolidation, c_v, in the range of 0.01 to 0.10 cm^2/sec. For comparison, low to high plasticity clays typically have coefficients of consolidation in the range of 0.001 to 0.0001 cm^2/sec.

The one-dimensional coefficient of secondary compression, C_α, governs the magnitude of secondary compression or drained creep deformations which occur after consolidation is complete. For the Widows Creek FGD gypsum-fly ash waste, the coefficient of secondary compression was found to vary from 0.09% to 0.23% with an overall average of 0.15%. This value is relatively small, and hence the effects of secondary compression are not projected to be significant.

PERMEABILITY CHARACTERISTICS

The void ratio, e, versus coefficient of permeability, k, relationship developed from laboratory permeability tests on resedimented and undisturbed samples of Widows Creek FGD gypsum-fly ash waste is presented in Figure 12. As shown, over the range of void ratio (or solids contents) of the test specimens, the coefficient of permeability is generally insensitive to void ratio and varies within the range of $5\text{x}10^{-5}$ to $5\text{x}10^{-4}$ cm/sec. At a void ratio of 1.1, typical of the void ratio at a depth of about 1 foot, the coefficient of permeability is on the order of $2\text{x}10^{-4}$ cm/sec. At a void ratio of 0.8, coresponding to an effective consolidation stress of about 4.0 kg/cm^2 (Figure 11) typical of the effective stress at the base of a 150-foot high stack, the coefficient of permeability is on the order of $1\text{x}10^{-4}$ cm/sec. Accordingly, the coefficient of permeability of FGD gypsum-fly ash waste disposed of within a stack is not projected to vary significantly.

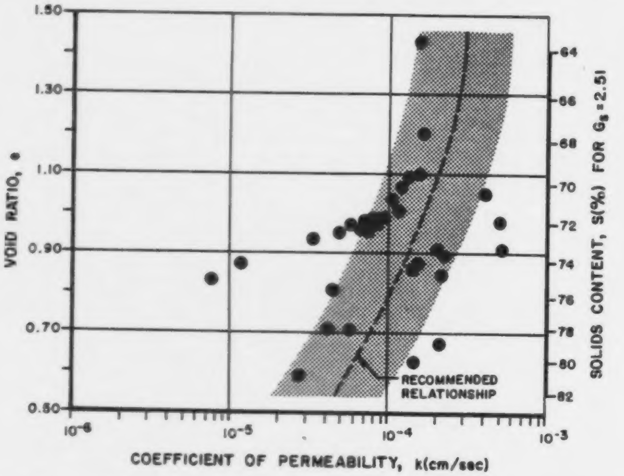


Figure 12. Void Ratio Versus Coefficient of Permeability.

STRESS-STRAIN-STRENGTH BEHAVIOR

The relationship for Widows Creek FGD gypsum-fly ash waste between preshear dry density and peak angle of internal friction determined in drained ($\bar{\phi}_d$) and undrained ($\bar{\phi}$) shear for zero effective cohesion ($\bar{c}$) is presented in Figure 13. As shown, the angle of internal friction increases from 40° at dry densities of 72.5 to 80.0 lb/ft^3 to 49° at a dry density of 99.0 lb/ft^3. For the range of dry densities expected to exist within a stack, the lower angle of internal friction of 40° with zero effective cohesion is projected for Widows Creek gypsum-fly ash waste disposed of via wet stacking.

Stress-strain curves for Widows Creek FGD gypsum-fly ash waste from three consolidated drained triaxial tests performed on undisturbed samples recovered from the pilot-scale stack are presented in Figure 14. The test samples displayed pre-shear solids contents, S, of 71.2% to 74.1%, dry densities, γ_d, of 78.2 lb/ft^3 to 84.0 lb/ft^3 and were isotropically consolidated under effective confining stresses, $\bar{\sigma}_c$, of 0.5 to 3.0 kg/cm^2. The samples display similar behavior mobilizing peak strengths at axial strains, ϵ, of 6.7% to 11.5% and displaying slight strain softening behavior or decreases in strength with additional strain. Volume change measurements during drained shear indicate that the samples initially compressed (i.e., positive volume change), prior to a slight increase in volume at large strains.

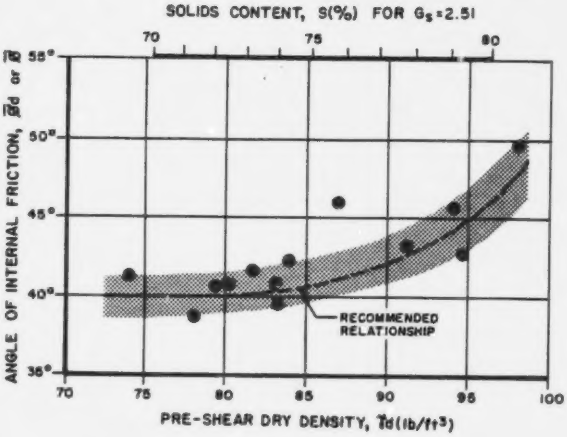


Figure 13. Dry Density Versus Angle of Internal Friction.

WET STACKING OF GYPSUM IN THE PHOSPHATE FERTILIZER INDUSTRY

Wet stacking of by-product gypsum, i.e., phosphogypsum, has been practiced by the phosphate fertilizer industry for more than 25 years. In Florida alone, more than 21 million tons of phosphogypsum are disposed of annually using the wet stacking method. The resulting gypsum stacks are typically large (50 to 300 acres), structurally stable stockpiles reaching heights greater than 100 feet. A gypsum stack located near Bartow, Florida is shown in Figure 15.

DESIGN CONSIDERATIONS

Wet stacking as performed by the phosphate fertilizer industry uses the upstream method of construction. In this method, illustrated in Figure 16, an earthen starter dike is first constructed to form a sedimentation pond and stacking area. Gypsum is pumped to the sedimentation pond in slurry form, usually at 15 to 20 percent solids, and allowed to settle and drain to approximately 60 to 70 percent solids. Process water is decanted and returned to the plant. Once sufficient gypsum sediments within the pond, the gypsum is excavated with a dragline to raise the perimeter dikes of the stack. The cast gypsum is then shaped to form a road on the crest of the dike using a bulldozer. The process of sedimentation, excavation, and raising of the perimeter dikes continues on a regular basis during the active life of the stack.

Using the upstream method of construction, some gypsum stacks have reached heights exceeding 100 feet with slopes as steep as 1.5 horizontal to 1.0

vertical which is approximately the angle of repose of phosphogypsum. The steep slopes result from casting the gypsum with a dragline and allowing some gypsum to roll down the outside of the stack thus eliminating shaping. Consequently, the outside slopes of some stacks have a factor of safety very close to unity and from a conventional geotechnical engineering point of view, failures of gypsum stacks do occur. Fortunately, gypsum is a very forgiving material, and unlike many mine tailings, gypsum does not readily flow. Therefore, the consequences of these failures are usually not dramatic.

In plan, the gypsum stack is usually a square or a rectangle. A typical layout is shown in Figure 17. The required area of the gypsum stack is generally selected to allow raising the perimeter dikes approximately 5 to 8 feet per year.

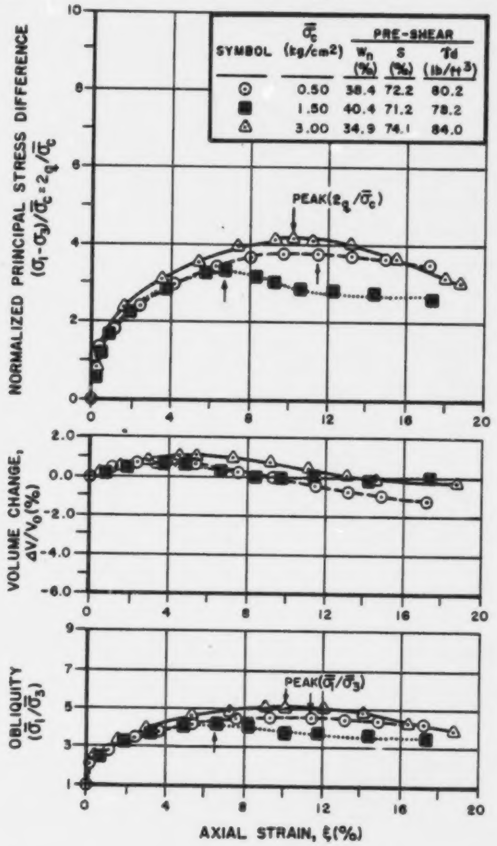


Figure 14. Drained Stress-Strain Behavior of Sedimented FGD Gypsum Fly Ash Waste.

OPERATIONAL CONSIDERATIONS

Two important considerations in the operation of a gypsum stack are the retention of fine-grained gypsum within the stack and the characteristics of the gypsum deposited around the periphery of the stack.

If sufficient retention time is not provided within the ponded area on top of the stack, complete clarification will not occur and the finer particles of gypsum will discharge through the spillway and sediment within the return ditch.

Although finer-grained gypsum can be cast to raise the perimeter dikes, the lower permeability and higher compressibility does not allow drainage to occur as quickly as with coarser-grained gypsum. Consequently, it is more difficult and time-consuming to use the finer material to raise the stack. For this reason, it is desirable to deposit the coarser material around the entire periphery of the stack.

Figure 15. Phosphate Fertilizer Plant Gypsum Stack Near Bartow, Florida.

It is possible to operate a gypsum stack so the coarser material is deposited around the perimeter of the stack while simultaneously providing sufficient retention time for the finer particles to settle within the interior of the stack. This is accomplished by using an elevated ditch to carry the gypsum around the periphery of the stack and to create a ponded area within the interior of the stack. This concept is shown schematically on Figure 17.

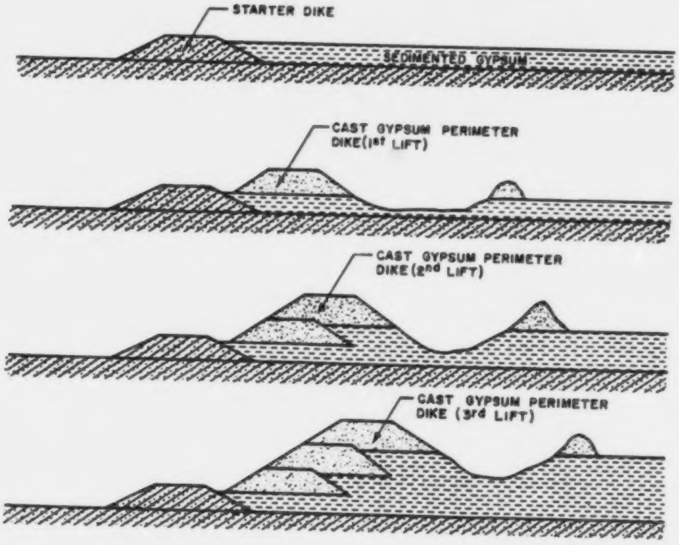

STARTER DIKE
SEDIMENTED GYPSUM
CAST GYPSUM PERIMETER DIKE(1st LIFT)
CAST GYPSUM PERIMETER DIKE(2nd LIFT)
CAST GYPSUM PERIMETER DIKE (3rd LIFT)

Figure 16. Upstream Method of Gypsum Stack Construction.

It is also desirable to divide the gypsum stack into at least two ponds. This allows one pond and its associated rim ditch to drain while the other pond is in use. The dragline can then excavate gypsum from the inactive rim ditch to raise the perimeter dike. It is often possible, however, using a dragline, to raise the stack without draining the gypsum.

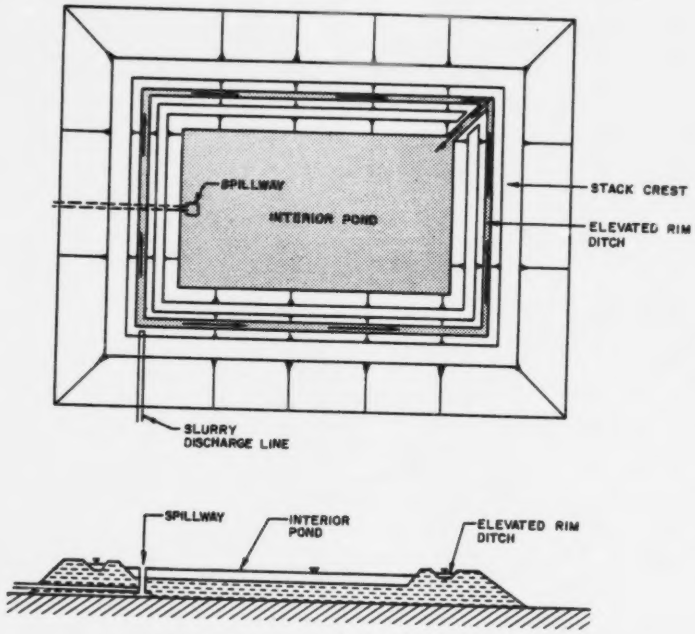

SPILLWAY
INTERIOR POND
STACK CREST
ELEVATED RIM DITCH
SLURRY DISCHARGE LINE
SPILLWAY
INTERIOR POND
ELEVATED RIM DITCH

Figure 17. Typical Gypsum Stack Layout Illustrating Elevated Rim Ditch.

Figure 18 illustrates the seepage pattern within a gypsum stack on an impervious foundation or liner. Under these conditions, the gypsum stack slope below the spring line is subjected to seepage forces. Under these conditions, sloughing of the outside slope of the gypsum stack may occur unless relatively flat slopes or internal drains are used.

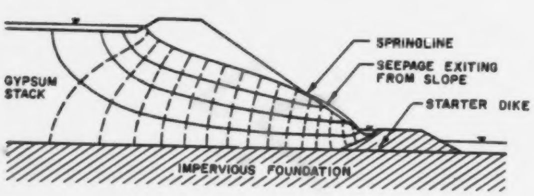

SPRINGLINE
SEEPAGE EXITING FROM SLOPE
GYPSUM STACK
STARTER DIKE
IMPERVIOUS FOUNDATION

Figure 18. Seepage Pattern Through a Gypsum Stack on an Impervious Foundation.

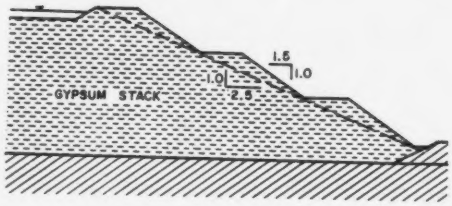

1.5
1.0
1.0
2.5
GYPSUM STACK

Figure 19. Flattening Gypsum Stack Slopes by Benching.

It is not common to provide erosion protection on the outside slope of a gypsum stack. Experience indicates there is essentially no erosion of the slope from rainfall, and dusting has not been a significant problem. If long-term maintenance and reclamation require the slopes be grassed, it may be expedient to flatten the slopes to 2.5 horizontal to 1.0 vertical or flatter as the stack is raised. These flatter slopes will generally hold topsoil cover and can be maintained using conventional equipment.

Gypsum stacks within the phosphate fertilizer industry are generally built as steeply as possible so that the storage capacity of the stack is maximized. If an average slope flatter than the angle of repose is required for stability, the perimeter dikes are generally offset from the outer perimeter as shown in Figure 19 to form benches in the slope. The excess material can be removed with a dragline to provide a uniform slope.

The wet stacking method of waste disposal was also demonstrated for FGD gypsum during operation of the Chiyoda Thoroughbred 121 Scrubber at the Scholz Electric Generating Station of Gulf Power Company in Sneads, Florida (6). A pilot scale gypsum stack similar to those used in the phosphate industry was successfully constructed (Figure 20) at Plant Scholz using the essentially pure gypsum produced by the CT 121 process.

EVALUATION OF WET STACKING DISPOSAL OF WIDOWS CREEK FGD GYPSUM-FLY ASH WASTE

Gypsum-fly ash waste was initially discharged into a 1.0-acre disposal area at Widows Creek on November 15, 1982 and continued intermittently through August 4, 1983. During this time, the scrubber was operational for 70 days and sufficient waste was provided to raise a 0.3-acre stack to a height of about 10 to 12 feet. The venturi bleed stream solids content varied considerably during operation of the scrubber from a minimum of 1.6% to a maximum of 15.7%, with an average of 8.1%. Based on an average venturi bleed stream solids content of 8.1%, average venturi bleed stream flow rate of 100 gal/min and 70 days of scrubber operation, an estimated 3,600 tons of waste were deposited within the disposal area. Waste deposited within the disposal area was characterized by approximately equal portions of gypsum and fly ash with average constituent contents of 42% gypsum, 43% fly ash, 13% unreacted limestone and 0.7% unoxidized calcium sulfite.

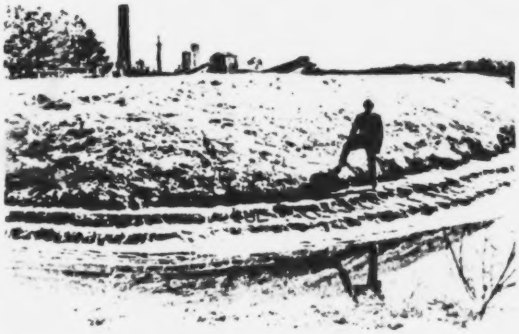


Figure 20. CT-121 FGD Gypsum Stack.

OPERATIONAL EXPERIENCE AND IMPLICATIONS

Photographs of stacking operations are presented in Figures 21 and 22. Excavation of gypsum-fly ash waste and casting onto the surface of sedimented waste to form the perimeter dike is illustrated in Figure 21. As shown in this photograph, relatively coarser and drier gypsum-fly ash waste near the slurry inlet line was easily cast. Relatively wetter gypsum-fly ash waste further away from the slurry inlet line did not cast well, and tended to slough and "run" after casting (Figure 22).

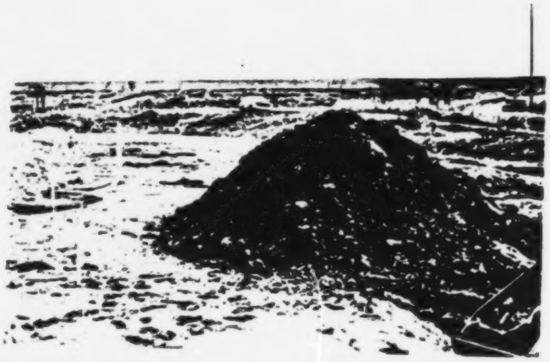


Figure 21. Casting of Gypsum-Fly Ash Waste.

The cast perimeter dike was not trafficable with a dozer until after several days of drying and dewatering. After additional drying and dewatering, rehandling and recasting, and grading by a dozer the perimeter dike was successfully constructed as shown in Figures 23 and 24. Note that the cast perimeter dike slopes are relatively steep, generally 1.0 horizontal to 1.0 vertical.

During construction of the stack, the pond was drained and a drying/dewatering period of 7 to 53 days allowed prior to stacking. Even with the drying/dewatering period, however, only coarser and drier material near the inlet to the stack was easily cast. Sedimented gypsum-fly ash waste further from the inlet was difficult to cast and required considerable rehandling due to the tendency to slough and "run" after casting.

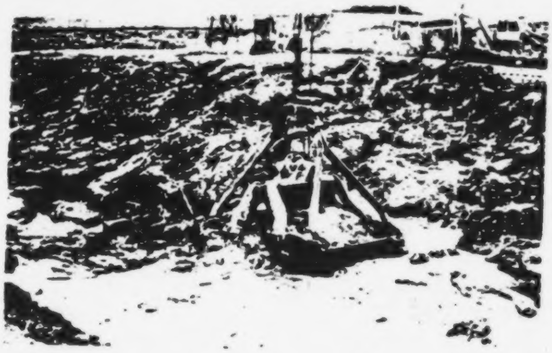


Figure 22. Casting of "Wet" Gypsum Fly Ash Waste.

At the end of the first few days of construction the trafficability of the cast dikes was generally poor. In some areas, the cast dike was not capable of supporting a small dozer. With continued rehandling and drying/dewatering, the cast dikes were satisfactorily constructed (Figure 24). Based on the results of in situ moisture content and density determinations, it appears that drying/dewatering of the gypsum-fly ash waste to a solids content of 80% is necessary before the material can be easily cast and shaped into a trafficable perimeter dike.

The slopes of the cast perimeter dike after stacking were steep, generally 1.0

horizontal to 1.0 vertical. Below the line of seepage the slopes were flat and sloughing of the cast slope immediately above the line of seepage was observed. The use of relatively flat slopes at the toe of the stack or an internal drain may be necessary to prevent similar conditions in a full-scale facility.

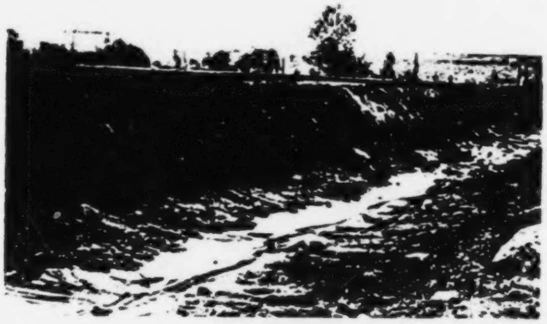


Figure 23. Completed Cast Gypsum Fly Ash Perimeter Dike.

In a full-scale wet stacking disposal facility some of the handling problems encountered during the demonstration project may be minimized by the use of an elevated rim ditch along the perimeter of the stack. Rehandling of the waste during casting, however, may still be required. The ability to allow drying/dewatering periods of 30 days in the rim ditch and/or pond prior to raising the stack should be considered in the design of a stack. Further, the stacking characteristics are sensitive to the fly ash and calcium sulfite content. Accordingly, gypsum-fly ash waste should be produced with a minimum amount of fly ash and calcium sulfite. Based on experience at Widows Creek, fly ash contents should be maintained less than 40% and calcium sulfite contents less than 1%. Improvement in the stackability of the gypsum-fly ash waste will result if the fly ash and calcium sulfite contents are minimized and the gypsum content maximized.

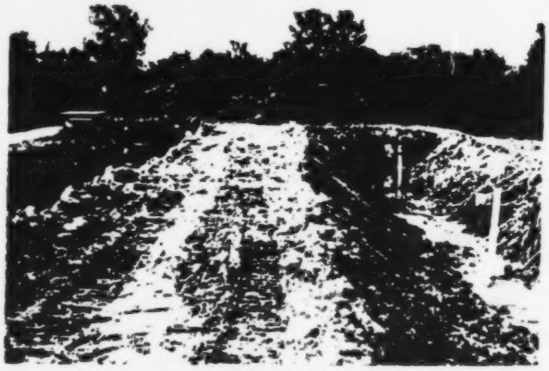


Figure 24. Completed Cast Gypsum Fly Ash Perimeter Dike.

The work described in this paper was prepared by Ardaman & Associates, Inc. (Ardaman), and the Tennessee Valley Authority (TVA). Neither Ardaman, TVA, nor any persons acting on their behalf: make any warranty or representation, expressed or implied, with respect to the use of any information contained in this paper, or that the use of any information, apparatus, method, or process disclosed in this paper may not infringe privately owned rights; or assume any liabilities with respect to the use of, or for damages resulting from the use of, any information, apparatus, method, or process disclosed in this paper. This paper does not necessarily reflect the views and policies of TVA.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

REFERENCES

1. Patterson, J.G., Munson, G.T., Runyan, R.A., Veitch, J.D., and Wells, R.L., "Forced-Oxidation Demonstration at the TVA Widows Creek Unit 8 FGD System." Interagency Energy/Environment R&D Program Report. TVA/OP/EDT-82/15, 1982.

2. Wells, W.L., Runyan, R.A., Munson, G.T., Moore, N.D., and Crowe, J.L., "Limestone Counter Current Scrubber Optimization Test and Forced Oxidation of Limestone Scrubber Sludge at TVA's Widows Creek Unit 8 Steam Plant." Presented at the 42nd Annual American Power Conference, Chicago, Illinois, 1980.

3. Wells, W.L., Muirhead, W.B., and Buckner, J.H., "TVA's Experience with Limestone Scrubbers at the 550-MW Widows Creek Unit 8." Presented at the 40th Annual American Power Conference, Chicago, Illinois, 1978.

4. Garlanger, J.E. and Ingra, T.S., "Evaluation of Engineering Properties and Wet Stacking Disposal of Widows Creek FGD Gypsum-Fly Ash Waste." Prepared for Tennessee Valley Authority, Office of Power, Division of Energy Demonstrations and Technology, Contract TV-58330A, 1983.

5. Utley, D.L. and Jones, B.F., "Field and Laboratory Study of Gypsum Landfilling at TVA's Widow Creek Station." Radian Corporation, 1980.

6. Hagerty, D.J. and Ullrich, C.R., "Interim Report-Physical Characterization of FGC Wastes." Report to Arthur D. Little, Inc., EPA Contract 68-02-2654, 1980.

7. Garlanger, J.E. and Ingra, T.S., "Evaluation of Chiyoda Thoroughbred 121 FGD Process and Gypsum Stacking." Volume 3, Electric Power Research Institute, Research Project RP-536-3, Final Report CS-1579, 1980.

8. Kingston, W.H., Wright, C.E., Jr., Breault, R.W., Mohn, N.C. and Mahavizadeh, G.B., "Demonstration of Forced Oxidation at TVA's Widows Creek Unit 8 FGD System." Final Report, C-E Environmental Systems, Inc., 1979.

SOLIDS HANDLING AND DEWATERING OPTIMIZATION AT COLORADO UTE'S CRAIG STATION

by: O.W. (Buddy) Hargrove
and David R. Owens
Radian Corporation
Austin, Texas 78766

Dorothy A. Stewart
Electric Power Research Institute
Palo Alto, California 94303

Jim Renner
Colorado Ute
Montrose, Colorado 81401

ABSTRACT

Units 1 and 2 at Colorado Ute's Craig Station include wet limestone scrubbers for SO_2 emissions control. Because the SO_2 concentration in the gas is very low, the solids formed are completely oxidized. Normally, gypsum produced in a limestone FGD system dewaters and handles relatively well. However, the centrifuge product at Craig fluidizes when shear forces are applied and, as a result, cannot be transported in an open truck.

EPRI, Radian, and Colorado Ute have been investigating the poor handling properties of these solids over the past 3 years. The fluid properties of the solids are a result of the fine crystals in the product cake which may be caused by a number of different factors. This paper discusses possible causes of the cake properties and methods to optimize the performance of the dewatering equipment. A comparison of full-scale centrifuge and pilot-scale vacuum filter results is included.

INTRODUCTION

The Craig Generating Station of Colorado-Ute Electric Association participated in an Electric Power Research Institute research project to define and solve chemical process problems in flue gas desulfurization (FGD) wet scrubbing systems. The problems at Craig were mist eliminator scaling and poor handling properties of the waste solids from the system.

Initial investigation to define the cause of these problems and to offer solutions has been reported in EPRI Report CS-3796, Field Investigation of FGD System Chemistry. However, after making the recommended changes in the FGD system, Craig continued to produce solids that were difficult to handle. Further work was done at this site by EPRI and Radian, Inc., as contractor to determine if a more easily handled product could be obtained.

BACKGROUND

The Craig Power plant is owned by a consortium of utilities and operated by Colorado-Ute. Units 1 and 2 are 400 MW capacity designed to burn 0.45 percent sulfur subbituminous coal mined on site. The flue gas is first treated by electrostatic precipitators for collection of fly ash, then by limestone scrubbers for SO_2 removal. Each unit is equipped with 4 spray towers with 3 designed to operate and 1 as a spare.

Figure 1 shows a schematic of one scrubber module. The L/G is about 60 gal/1000 acf with both pumps running. The reaction tanks are large enough to provide a solid phase residence time of several days with the low SO_2 concentrations in the flue gas. Limestone is added to maintain pH for SO_2 removal.

The primary differences between this flow scheme and that of many other limestone FGD systems center around the process hydroclones and the use of centrifuges for dewatering. A slip stream from the absorber feed stream is routed to the hydroclones. The underflow is sent back to the reaction tank while the overflow feeds a wash tank. Liquor from this wash tank was originally used to wash the interface tray and the mist eliminator. A modification has been made to wash the mist eliminator with cooling tower blowdown while the interface tray wash still comes from the wash tank. The overflow from the wash tank is routed to the baffled overflow area in the reaction tank. The reaction tank overflow feeds the thickener, and the thickener underflow is sent to one of three centrifuges for final dewatering.

Since initial operation of these units, sludge handling and disposal have been an operating inconvenience. In spite of sludge solids concentrations of greater than 80 weight percent, the cake is quite fluid. Originally, Colorado Ute wanted to backfill the coal mine with the sludge, but the open ash trucks could not be used without much of the sludge flowing out as the trucks negotiated the grade back to the mine. Freezing during the winter months was also a problem while using the ash trucks. Cement trucks were leased as an interim solution, but these trucks cannot handle the steep grade to the mine, so the sludge is currently disposed of in an on-site landfill.

In order to assist Colorado Ute with these problems, EPRI and Radian conducted sampling and data gathering trips to characterize the process and to identify possible underlying causes of the problems. From these initial trips, it was discovered that fine solids particles were being sent to the thickener and large particles were being retained in the reaction tank. The reaction tank crystals had rounded edges, indicating that the solids were remaining in the scrubbing loop long enough to show significant levels of attrition. The hydroclone was suspected as the primary cause of the fine particles entering the thickener, but the relatively quiescent baffled reaction overflow could also be contributing to particle classification.

A recommendation was made to Colorado Ute to reroute the hydroclone (wash tank) overflow stream to the center of the reaction tank rather than

Reproduced from
best available copy.

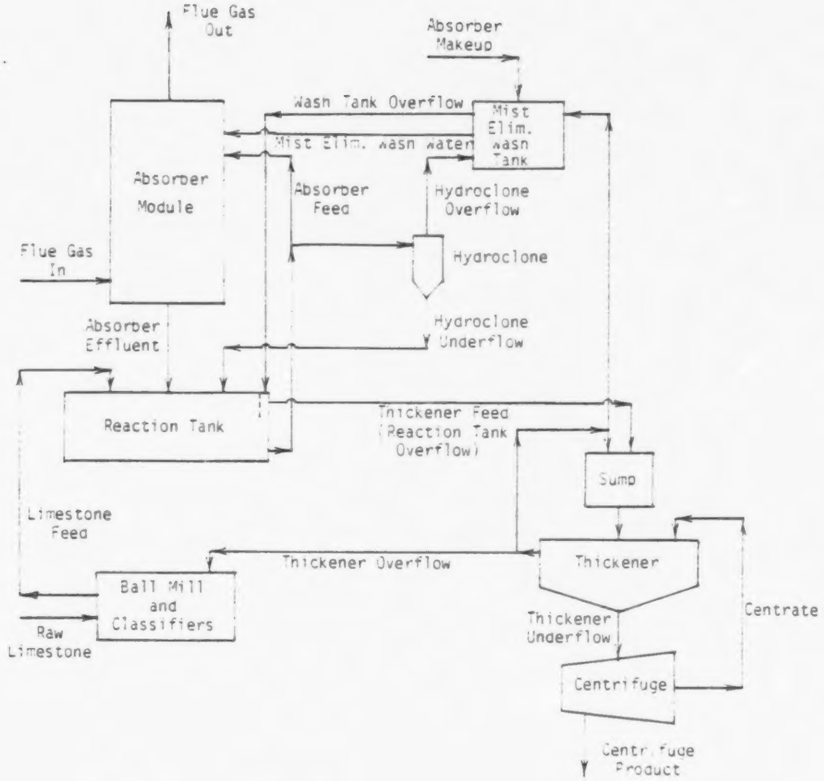

Flue Gas Out
Absorber Makeup
Wash Tank Overflow
Mist Elim. Wash Tank
Mist Elim. Wash Water
Absorber Module
Absorber Feed
Hydroclone Overflow
Flue Gas In
Hydroclone
Absorber Effluent
Hydroclone Underflow
Reaction Tank
Thickener Feed (Reaction Tank Overflow)
Sump
Limestone Feed
Thickener Overflow
Thickener
Ball Mill and Classifiers
Raw Limestone
Thickener Underflow
Centrate
Centrifuge
Centrifuge Product

Figure 1. Simplified flow diagram of the Craig FGD System.

into the baffled area. This should reduce the amount of fine material overflowing the reaction tank and entering the thickener. Modifications were made to the absorber modules on Unit 2, and both units were operated for several weeks to observe the effect of the modification on sludge properties. The particle size of the thickener feed solids was noticeably increased in Unit 2, and improved filtering and handling properties were noted in laboratory tests. There was an initial improvement in the centrifuge operation, but the Unit 2 solids were still somewhat fluid. Over a period of several months, Colorado Ute personnel noticed that the amount of fly ash entering the FGD system seemed to have a more dramatic effect on the solids properties than did the revision in the process.

Changes were made in the centrifuge settings to increase the solids content, but the changes in the dam settings seemed to have little effect on the centrifuge cake. It was also found that the centrifuge was classifying particles and returning fine particles to the thickener in the centrate stream. This results in smaller particles in the thickener underflow than in the thickener feed stream and contributes to the cake handling problems.

As a result of this sequence of events, a program to evaluate filtering as an option for Craig was formulated.

TEST PROGRAM

The difficulties associated with dewatering the solids by centrifuge were well documented by Colorado Ute and EPRI/Radian during the initial sampling trips. Because of indications that the initial process change, which minimized concentration of fine crystals in the thickener feed stream, had not been completely successful in improving the solids dewatering, a pilot and bench filter program was undertaken. The objectives of this program were to:

- Determine if the Craig solids could be effectively dewatered by vacuum filtration;
- Determine if process changes made on the Unit 2 scrubbers had had a significant effect on solids properties;
- Propose additional changes in the process configuration which might result in improved solids characteristics; and
- Compare the performance of the full-scale centrifuge to that of the pilot vacuum filters.

Unfortunately, the maintenance condition of the centrifuges was such that a meaningful comparison with vacuum filtration could not be made.

In order to document the dewatering properties of the Craig solids using vacuum filtration, Colorado-Ute leased an Eimco bottom feed rotary drum vacuum filter (see Figure 2) and an Eimco extractor (horizontal belt vacuum filter, Figure 3). Both filters came equipped with a nylon cloth,

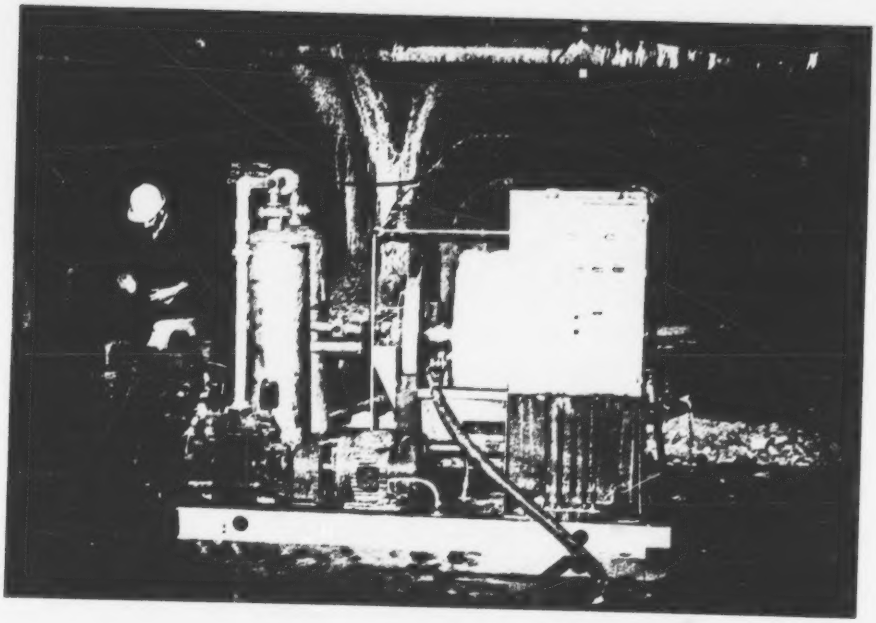


Figure 2. Picture of drum filter

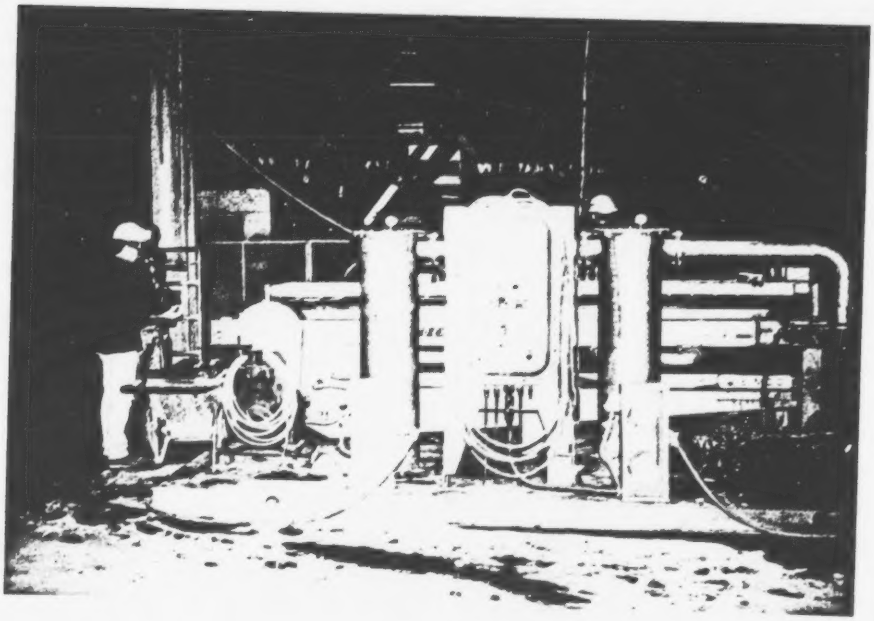


Figure 3. Picture of belt filter

NY-529F, which is a multi-filament yarn with a 2 x 2 twill weave, 144 x 54 threads per inch, and has a design air flow rate of 40 cubic feet per minute per square foot of cloth surface. The rotary drum cloth area was 10 square feet while the extractor had 10.5 square feet of cloth filtering area. The extractor was equipped with a vacuum pump designed to pull a 20-inch Hg vacuum at 38 cubic feet per minute air flow per square foot of cloth area. The rotary filter's vacuum pump was designed to pull the same vacuum at 11 cfm air flow/ft^2 of cloth.

The extractor came equipped with wash sprays to clean both sides of the cloth after the solids had discharged and the cloth had released from the belt. This minimized the effect of the fine solids adhering to the cloth and reducing the filtering rate (blinding) on subsequent passes. Initially, the rotary filter did not have any cloth washing provisions. However, since the Craig solids tend to blind the NY-529F cloth, one of the cake washing nozzles was relocated to wash the front side of the cloth after the solids had been discharged. The back side of the cloth could not be washed since the cloth did not release from the drum.

The pilot filter testing was conducted from November 14 to December 12, 1984. The streams tested on the drum filter included thickener underflow from both units and hydroclone underflow from Module 1B. The horizontal belt was operated with thickener underflow from Unit 1 and with hydroclone underflow from Module 1B. For several days during the test program, the horizontal belt filter was used to control the solids concentration in Module 1B by filtering hydroclone underflow. During the test program, the centrifuges dewatered thickener underflow from both units using normal procedures. In addition, hydroclone underflow was sent to the Unit 1 thickener for several days to simulate operation with hydroclone underflow as the feed to the centrifuge.

During the pilot test program, the bench-scale filter leaf set up was used to test other streams such as Module 2A and 1B recycle streams and Module 2A hydroclone underflow stream. A schematic of a top feed filter leaf apparatus is shown in Figure 4. In order to analyze the performance of a given slurry on a horizontal belt, aliquots of the slurry are poured into the apparatus to be filtered. Different volumes and solids concentrations are used to document the performance over a range of possible operating conditions. The time required for the liquid to disappear from the cake surface is the form time. This apparatus can also be used to determine the dry time required for a given slurry.

In addition to the filter leaf set up, several other analytical procedures were used to document the characteristics of the Craig solids including:

- solids concentration (weight percent);
- particle size distribution (wet sieving and coulter counter);

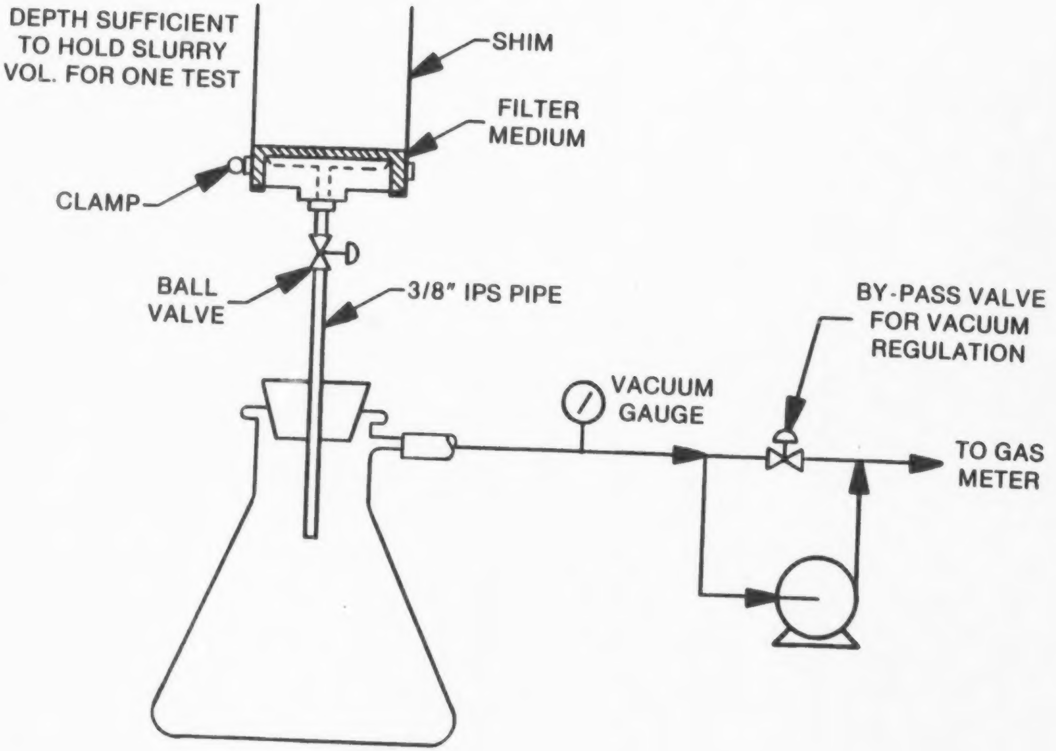

DEPTH SUFFICIENT
TO HOLD SLURRY
VOL. FOR ONE TEST
SHIM
FILTER
MEDIUM
CLAMP
BALL
VALVE
3/8" IPS PIPE
VACUUM
GAUGE
BY-PASS VALVE
FOR VACUUM
REGULATION
TO GAS
METER

Figure 4. Top-feed filter leaf test arrangement.

- crystal habit (photomicrographs and scanning electron micrographs); and

- solids composition (infrared spectroscopy, atomic absorption, ion chromatography).

RESULTS

Overall, the results of the program indicated that the hydroclone underflow stream dewaters much more efficiently on the vacuum filters than does thickener underflow. Filtering properties of the thickener underflow solids changed significantly during the test program, probably because of variations in the amount of fly ash present in the slurry. The centrifuge product also varied considerably with time, but in all cases the product remained somewhat fluid in nature.

PILOT FILTER RESULTS

Table 1 compares operation of the drum filter with different slurry feed streams. Thickener underflow from Unit 1 severely blinded the cloth during the early phases of the program. However, Unit 1 had been operating for less than one week after its fall maintenance outage. Laboratory results showed that the filtering properties of this stream did improve through the course of the program, but not to the point where another drum filter run was attempted.

The thickener underflow stream from Unit 2 exhibited much better filtering properties during the initial phases of the program than did thickener underflow from Unit 1. A 2-inch cake from the rotary drum was achieved on November 15, and a filtering rate of 790 pounds per hour was measured. Some blinding of the cloth occurred, and the solids fluidized when shear was applied. The cloth was washed and another run was made which produced a thinner cake. The cake solids increased from 62 to 76 percent, and, while the solids still fluidized upon shearing, more shear was required to effect the same result.

The properties of the Unit 2 solids changed dramatically as the test program proceeded. The rotary drum filter test conducted on December 6 shows that the filtration rate was reduced by a factor of 3. The change in operation of the centrifuge was even more apparent with an extremely fluid centrifuge product produced during the early December period. Colorado Ute personnel predicted this result based on low unit load for several days prior to December 6. The hot-side ESP is not as efficient at low loads as it is at higher loads, and evidently higher than normal levels of fly ash were being removed in the scrubbers.

The hydroclone underflow from Module 1B showed the best filtering properties of all of the runs made with the rotary drum filter. Table 1 shows that a filtering rate of 1570 pounds per hour was achieved on December 6. The cake solids did not fluidize when shear force was applied, and the solids were of a sandy quality.

TABLE 1. SUMMARY OF DRUM FILTER RESULTS

Stream	Unit 1 Thickener Underflow	Unit 2 Thickener Underflow	Unit 2 Thickener Underflow	Module 1B Hydroclone Underflow
Date	11/14/84	11/15/84	12/06/84	12/06/84
Feed Solids, wt %	34	47	38	41
Cake Solids, wt %		62	80	78
Cake rate, lb/hr		790	258	1570
Comments	severely blinded cloth	2" cake, then blinded, solids fluidized upon applying shear	3/8" cake, solid fluidized upon applying shear	2" cake, solids did not fluidize upon applying shear

The horizontal belt filter tests showed that the hydroclone underflow was a consistently better filtering material than was thickener underflow. Table 2 summarizes the results of the belt filter tests. The Unit 1 thickener underflow, which completely blinded the cloth on the drum filter, was filtered somewhat more successfully on the belt filter. A half-inch cake was produced although the filtration rate was extremely low and there was a tendency for the cloth to blind even with the cloth washing sprays in service. On the other hand, the hydroclone underflow stream dewatered to greater than 85 percent solids with 1 to 3 inch cakes typically achieved. A filtration rate of greater than 5000 pounds per hour was measured for the 10.5 square foot pilot extractor.

Since the hydroclone underflow exhibited such good filtering properties, Module 1B was isolated from the rest of the system, and the density in the module was controlled by filtering the hydroclone underflow. This test was conducted over a 7-day period (November 29 to December 6) with encouraging results. Throughout this period, the cake properties were consistently evaluated as acceptable for transportation by open truck. The solids did not fluidize upon application of manual shear force or during the short trip from the scrubber to the on-site landfill. While this test was encouraging, it was not possible to conduct the test for a sufficient length of time to completely achieve steady state with respect to solids properties. With the long solid-phase residence time at Craig, such a test should be conducted for a minimum of one month.

CENTRIFUGE-FILTER COMPARISON

Since the hydroclone underflow stream filtered so well, a direct comparison between the filter and the centrifuge was made using hydroclone underflow slurry. The only convenient means of delivering hydroclone underflow to the centrifuge was through the thickener so hydroclone underflow from Modules 1B and 1D was fed to the Unit 1 thickener for several days. While there appeared to be some improvement in the centrifuge cake based on visual inspection, the cake was still fluid, and the solids concentration was still between 60 and 70 weight percent. The dam setting was changed from 3.6 to 2.3 in an effort to increase the solids concentration and the cake handling properties. This change should have reduced the pond depth and increased the length of dry beach. The centrifuge cake properties did not significantly change with the change in dam setting. The cake remained fluid and less than 70 percent solids. Colorado-Ute personnel report considerable centrifuge wear, particularly on the tiles, and this may be limiting performance. As a result, this was not believed to be an adequate test of the capability of this centrifuge operating in good maintenance condition.

Simultaneously with the revised dam setting run on the centrifuge, a test with the Unit 1 thickener underflow was conducted on the drum filter. Good filtering properties were observed, and the product did not fluidize upon application of shear. A filtering rate of 860 pounds per hour was measured, and the cake thickness was 2 inches. Some blinding was also observed, but a drum designed for cloth release from the drum and cloth washing would help mitigate the blinding effects. A more open cloth weave would also be desirable.

TABLE 2. SUMMARY OF BELT FILTER RESULTS

Stream	Unit 1 Thickener Underflow	Module 1B Hydroclone Underflow	Module 1B Hydroclone Underflow	Module 1B Hydroclone Underflow
Date	11/14/84	11/14/84	11/30/84	12/03/84
Feed Solids, wt %	34	43	47	43
Cake Solids, wt %	75	93	83	85
Cake rate, lb/hr	-	-	3000	5700
Comments	1/2" cake; blinding; noted solids fluidize upon shearing	1-2" cake; solids handled well; no fluidization observed	1-3" cake; solids handled well; no fluidization observed	1-3" cake; solids handled well; no fluidization observed

BENCH-SCALE TEST RESULTS

Throughout the program, bench-scale filter tests were conducted to compare properties of different streams in the process. The filter leaf test results support the conclusion that the hydroclone underflow stream has better filtering properties than other streams within the process. Table 3 shows results of filter leaf tests performed on samples taken on December 5, 1984. Hydroclone underflow from Module 1B had the shortest form time (8 seconds) and highest form filtration rate (1300 pounds per hour per square foot). Hydroclone underflow from Module 2A had a slightly longer form time and a 30 percent slower form filtration rate, but the filtering properties were judged to be quite acceptable.

Module 1B recycle also exhibited fairly good filtering properties. The form time and filtration rate were similar to those of the 2A hydroclone underflow stream. However, Module 2A recycle slurry filtered much more slowly than did the 1B recycle stream. The 2A recycle stream form time was about 3 times slower (31 seconds) and the filtration rate (320 $lb/hr\text{-}ft^2$) was a third of that for 1B recycle.

Neither of the thickener underflow streams filtered well, although late in the program Unit 1 underflow filtered considerably faster than did Unit 2 thickener underflow. The form time for the Unit 2 thickener underflow stream was 45 times slower than the form time for the 1B hydroclone underflow stream. The form time for Unit 1's thickener underflow stream was 38 seconds which is almost 5 times slower than for 1B hydroclone underflow.

A major factor influencing the filtration rate is the amount of fine material in the solids, not the mean particle size. Table 4 shows that the mean particle size of the 1B recycle was 23 microns while the mean particle size for the Unit 2 thickener underflow was 37 microns. In spite of the 60 percent larger mean particle size, the thickener underflow stream filtered about 17 times slower than did the 1B recycle stream. Table 4 also shows the percentage of particles less than 5 microns in the minus 325 mesh fraction. Hydroclone underflow from Module 1B had no measurable amount of fines below 5 microns according to Coulter Counter results. This stream also exhibited the fastest filtering rate of the four streams shown. Unit 2 thickener underflow was the slowest filtering material and also had the greatest amount of fine material. However, note that the effect of the fines on the filtering is certainly not linear. For only 1 percent increase in the amount of material less than 5 microns, the filtering rate of Unit 2 thickener underflow decreases by a factor of 4 compared to the 2A recycle stream. These very fine particles evidently fill the volume between the larger solid particles and make dewatering the slurry more difficult.

Figure 5 shows scanning electron micrographs for the samples taken on December 5, 1984. The SEM's indicate that the particle size distribution below 5 microns is important in determining a slurry's filtering properties. Although 2A recycle and Unit 2 thickener underflow have about the same fraction of solid particles smaller than 5 microns, the thickener underflow stream has many more extremely fine particles, many of which appear to be less than 1 micron. Unfortunately, the Coulter Counter cannot differentiate

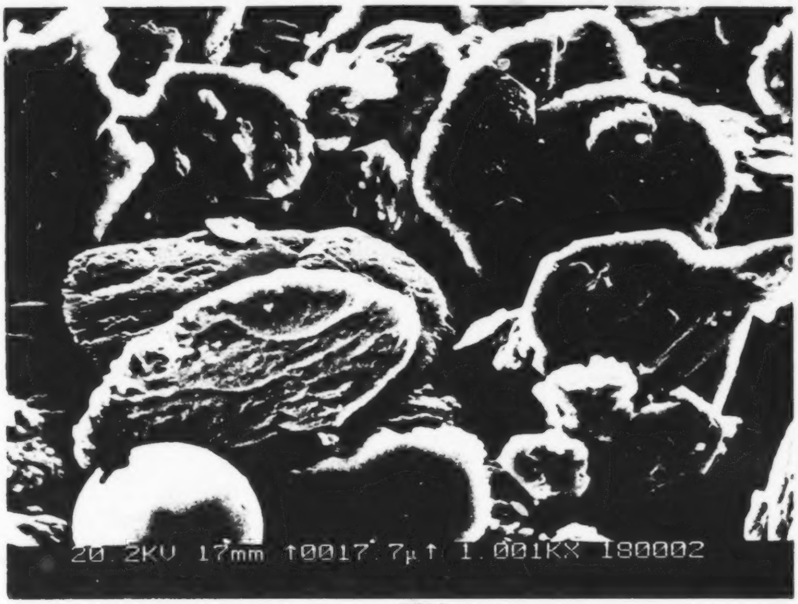
TABLE 3. SUMMARY OF FORM FILTRATION TEST RESULTS

Stream	Form Time, Sec.	Cake Solids, %	Form Filtration Rate (lb/hr-ft^3)
1B Recycle	10	75	1000
2A Recycle	31	75	320
1B Hydroclone Underflow	8	83	1300
2A Hydroclone Underflow	11	83	1000
Unit 1 Thickener Underflow	38	78	230
Unit 2 Thickener Underflow	360	75	59

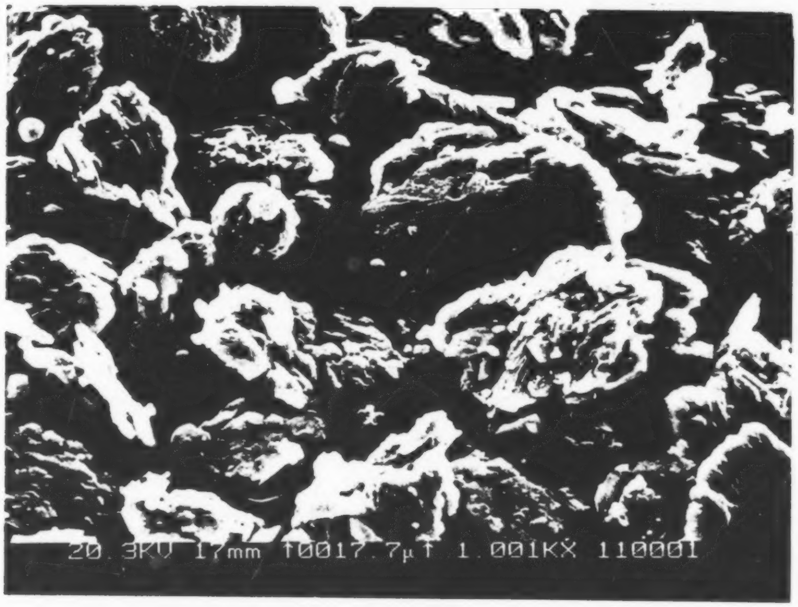
TABLE 4. EFFECT OF FINES ON FILTRATION RATE

Stream	Form Filtration Rate (lb/hr-ft^2)	Mean Particle Size (μ)	Percentage of Particles <5 Microns in -325 Mesh Fraction
1B Hydroclone Underflow	1300	40	0
1B Recycle	1000	23	4
2A Recycle	230	41	10
Unit 2 Thickener	59	37	11

Reproduced from
best available copy.

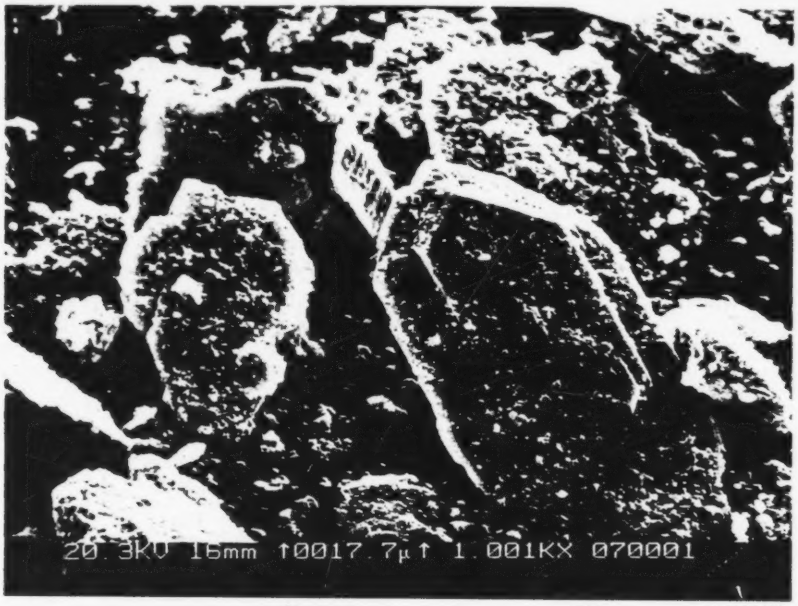

20.2KV 17mm ↑0017.7µ↑ 1.001KX I80002

1B Hydroclone Underflow

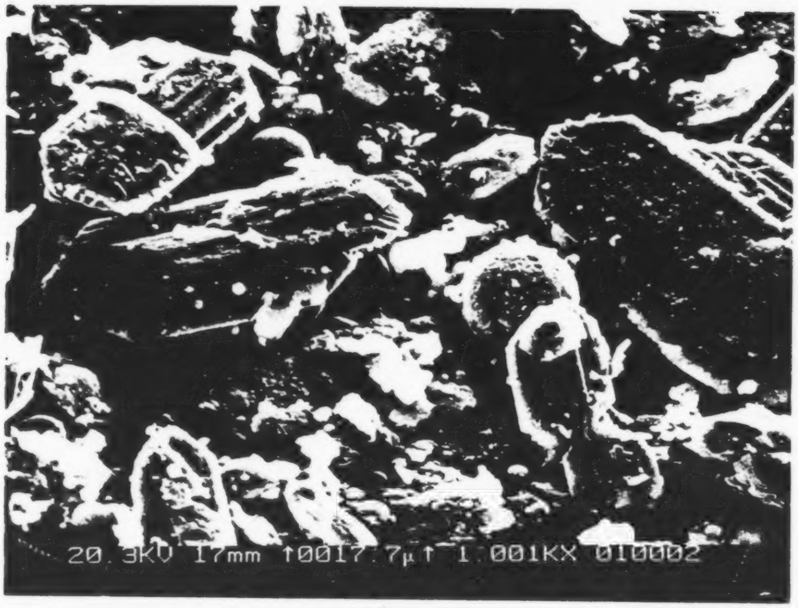

20.3KV 17mm ↑0017.7µ↑ 1.001KX 110001

1B Recycle

Figure 5. Scanning Electron Micrographs from 12/5/84


20.3KV 16mm ↑0017.7µ↑ 1.001KX 070001

Thickener Underflow Unit 2


20.3KV 17mm ↑0017.7µ↑ 1.001KX 010002

2A Recycle

Figure 5 (continued). Scanning Electron Micrographs from 12/5/84 between particles of 5 microns and 1 micron size with great accuracy, so the effect of fine particles on filtering could not be quantified as well as would be desired.

The concentration of fine particles in the thickener may be a direct result of the classification of fines by the centrifuge. In the Craig system, the centrate, with its higher concentration of fines, is returned to the thickener thereby increasing the fraction of fines in the thickener underflow. The thickener underflow pumps and the centrifuge may also be creating fines by attrition, but this could not be determined by direct measurement on one pass through the pumps or centrifuge.

Sedimentation tests were conducted on two of the thickener underflow samples and the fine (slow settling) fractions were analyzed by Inductively Coupled Argon Plasma Emission Spectroscopy (ICAP) and by infrared spectroscopy (IR). Relatively high concentrations of aluminum and silica indicate that fly ash inerts are a large portion of the fines. IR also shows that considerable amounts of calcium carbonate and gypsum are present.

One reason that the 1B recycle slurry had fewer fines than the 2A recycle slurry may have been the difference in slurry solids concentration. Module 1B was controlled at between 15 and 20 percent solids for several days prior to December 5, while Module 2A solids concentration remained between 2 and 5 percent. The longer solid phase residence time afforded by the higher density setpoint allows limestone to dissolve more completely and gypsum fines to increase in size. However, the largest difference between the two modules could be the amount of fine fly ash entering the scrubber.

FILTER SIZING

Based on the filter leaf test results, a rotary drum and a horizontal belt filter were sized for application at Craig. First, the solids production rate was estimated at 5700 pounds per hour per unit. If the filter is sized to operate 10 hours per day, the filtering capacity required is 13,700 pounds per hour per unit.

The specifications for the filter sizing are summarized in Table 5, and the results are summarized in Table 6. The filter size is controlled by the time required to dry the solids at Craig. Since the form time is much shorter than the dry time and the horizontal belt can be designed to accommodate this much better than a drum can, the horizontal belt is about 60 percent the size of the drum. The estimated uninstalled equipment cost for the horizontal belt is slightly less than that for the drum. Space and retrofit costs will decide the selection of the specific filter. However, if the retrofit costs are not too great, a horizontal belt would be recommended because of its flexibility.

Before a decision to choose a filter is made, investigation into limitations and possible modifications of the centrifuges should be conducted.

TABLE 5. SUMMARY OF FULL-SCALE FILTER DESIGN SPECIFICATIONS

Filtered Material	Hydroclone Underflow
Feed Solids Conc., wt %	35
Cloth	NY-529F
Cake Thickness, inches	7/8
Cake Solids Conc., wt %	81

TABLE 6. SUMMARY OF FILTER DESIGN RESULTS

	Horizontal Belt Filter	Rotary Drum Filter
Cloth Area, ft^2	120	200
Receiver Size	15" x 60"	15" x 60"
Moisture Trap Size	24" x 60"	24" x 60"
Filtrate Pump Size	5 hp	5 hp
Vacuum Pump Size	20 hp	40 hp
Equipment Costs, $1000 (not installed)	160	174

Note: Both systems are constructed of stainless steel with a local control panel and auxiliary platform.

CONCLUSIONS

- Craig's waste solids can be dewatered to a non-fluid product with either a drum or belt filter.
- The horizontal belt filter is more versatile and has a lower equipment cost than the drum filter for the Craig application.
- Centrifuge inefficiency may be due to design, operating, or maintenance limitations.
- The classification of particles in the centrifuge with the fines returning to the thickener in the centrate stream results in concentration of fines in the thickener. This increase in fines contributes to the poor dewatering properties of thickener underflow solids.
- Dewatering and handling properties appear to be a function of the amount of fine material (<5 microns).

RECOMMENDATIONS

- Evaluate improvement in centrifuge design and state of maintenance.
- Perform cost/benefit analysis of installing vacuum filtration.
- If economics are favorable for filters, conduct evaluation of hydroclone operation on long-term dewatering characteristics.
- If centrifuge operation is the optimum solution, reroute the centrate stream to the modules to minimize the concentration of fines in the thickener.
- Modify Unit 1 scrubbers to feed the thickener with either recycle slurry or with hydroclone underflow and not with hydroclone overflow.

EPRI RESEARCH ON MANAGEMENT OF WASTES FROM SO2 CONTROL PROCESSES

by: Dean M. Golden, P.E.
Electric Power Research Institute
Heat, Waste and Water Management Program
Palo Alto, California 94304

and

Russell H. Boyd, P.E.
Envirosphere Company
Norcross, Georgia 30092

ABSTRACT

The Electric Power Research Institute is keenly aware of the myriad of issues facing the electric utility industry in the area of controlling flue gas emissions from coal-fired power plants, particularly sulfur oxides. Virtually all facets of the question of how to control SO_2 emissions have been or are under consideration by the Institute in one or more research programs. In closing the loop on the question of SO_2 control technologies, however, EPRI recognized that the subject of management of solid wastes from the new control technologies had not been examined in depth.

The purpose of this paper is to provide a description of the research EPRI is conducting with regard to management of the solid waste materials from these new SO_2 control technologies. A discussion of the preliminary results of the research project also is presented. The research project focuses on spray drying, atmospheric fluidized bed combustion, limestone furnace addition, dry sodium addition, and advanced physical coal cleaning. Within the project are activities dealing with information surveys and assessments, waste characterization, leachate testing, conceptual designs, case studies, and utilization options. The project is comprehensive in scope and will be providing topical reports on various activities during the next three years.

EPRI RESEARCH ON MANAGEMENT OF WASTES FROM SO2 CONTROL PROCESSES

INTRODUCTION

The electric utility industry has incurred a significant burden in complying with various environmentally oriented laws and regulations, which essentially saw their beginning with the National Environmental Policy Act of 1969 (NEPA). The Clean Air Act Amendments of 1970 (CAA), and subsequent revisions, have had the most pronounced impact on site selection, permit application, preconstruction monitoring, design, construction and operation. In fact, air pollution control facilities had cost the industry an estimated additional $15.2 billion from 1969 through the end of 1982, over continuing with those practices in place in 1968. This additional cost does not include costs associated with the management of the waste materials generated from these facilities. Requirements at coal-fired power plants for more efficient particulate control facilities and new flue gas desulfurization (FGD) facilities have certainly increased the cost for managing the waste materials by adding to the quantity of material to be managed and altering its chemical and physical makeup. With coal meeting an even greater percentage of future consumer demand, the additional tonnages to be managed will continue increasing, with FGD wastes representing a significant portion of this increase.

This presentation addresses new SO_2 control technologies the electric utility industry is applying or considering to meet increased coal use, and accompanying environmental limitations. In particular, this presentation focuses on research the Electric Power Research Institute (EPRI) is conducting with regard to management of solid waste materials, or by-products, from these new SO_2 control technologies. This research is necessary because solid waste materials from each process have chemical and physical properties different from those of fly ash or scrubber sludge from conventional pulverized coal-fired boilers. These differences could require changes in typical waste management design and operation practices and could have a substantial impact on the cost of waste management and the overall economics of the process. This issue of solid waste management is particularly important to EPRI because the Institute is sponsoring extensive research and development dealing with SO_2 control technologies.

The remainder of this presentation provides, in order, a brief description of the systems being considered, objectives of the overall EPRI program in this area and this specific research project, selected results from a preliminary planning study and, finally, a summary. The size of the new research project limits the ability to provide detail in this presentation; however, the major components of the project and the expected results are provided.

SO_2 CONTROL SYSTEMS UNDER EXAMINATION FOR THIS PROJECT

A number of new processes controlling SO_2 emissions from electric utility power plants through the use of dry sorbents are in various stages of development. These processes include lime addition after the furnace (spray drying), atmospheric fluidized bed combustion (AFBC), limestone addition in a pulverized coal-fired furnace, and sodium compound addition after the furnace and before the particulate control device. Although each process is in a different stage of development, there is growing interest in the matter of solid waste management, which includes utilization, for all processes. A brief description of the systems included under this project is presented below, along with the rationale for including the process.

Spray Drying

In spray drying, a commercially available and applied technology, most often a lime slurry is sprayed as a fine mist into the flue gas in a reaction vessel, where the mist reacts with SO_2 and dries to a fine particulate. The flue gas and particulates then pass to a fabric filter or electrostatic precipitator where the reaction products and fly ash are separated from the flue gas stream and collected in hoppers for subsequent management. Spray drying is included in this study, even though it is a commercially available technology, because some questions remain concerning the solid waste from the process, particularly in the area of utilization. In addition, a careful examination of solid waste characterizations and management experiences at facilities with spray dryers should provide valuable information in developing recommendations for solid waste management of materials generated by the other lime based processes.

Atmospheric Fluidized Bed Combustion

For atmospheric fluidized bed combustion, crushed coal and limestone are mixed and held in suspension in the furnace by an upflow of combustion air. The coal burns in the bed and in the freeboard above the bed, while the limestone is calcined to form lime which reacts with SO_2. During combustion, a portion of the bed material is continuously withdrawn to maintain optimum bed conditions. A portion of the bed material, fly ash, SO_2 reaction products and unreacted lime are elutriated and typically carried by the combustion gases through the furnace to cyclones, where a substantial fraction of the particulate matter is collected and recycled to the bed. The remaining particulate matter is carried by the flue gas to a fabric filter or electrostatic precipitator where it is collected for subsequent management. The AFBC technology is an emerging alternative utilities are considering for SO_2 control.

Limestone Furnace Addition

Limestone furnace addition, a process that is in an early stage of development, accomplishes SO_2 control through direct dry sorbent injection (typically limestone or lime) into the furnace of a conventional pulverized coal-fired boiler. The reaction products from this process, together with fly ash, are carried by the flue gas to a fabric filter or electrostatic precipitator, where the particulate matter is collected for subsequent management.

Dry Sodium Addition

Sodium compound addition before a fabric filter, the additional SO_2 control process, is nearing the commercial stage of development. In early pilot studies nahcolite and trona, naturally occurring forms of sodium carbonate and bicarbonate, were pulverized and added to the flue gas ahead of a fabric filter. The SO_2 in the flue gas reacted with the sorbent in the duct and in the fabric filter where the sorbent and fly ash collected to form the filter cake. In commercial applications, the sorbent may be a man-made or naturally occurring form of sodium compound; but, in any case, the waste product will be a mixture of fly ash, reaction products and unreacted sodium compounds.

Advanced Physical Coal Cleaning

The processes discussed above are effective for controlling SO_2 emissions during and after the combustion of coal. An alternative action, in some situations, is to use a precombustion process such as coal cleaning to lower SO_2 emissions. Coal cleaning removes certain minerals inherent in the coal, mined rock, and pyritic material. The coal cleaning process is designed to yield a fuel with lower ash content, lower sulfur content and provide more heating value per ton. This means that high-ash, high-sulfur coals become more attractive because flue gas emissions can be lowered by using this process, and either eliminate or reduce the need for on-line SO_2 control equipment.

Reasons for Selecting Systems

Because of process variations, the end product of each SO_2 process may require new or modified handling, storage, transportation, disposal and utilization practices. And, in all cases, fly ash will be intimately mixed with reaction products producing a greater quantity of product to manage. In comparison with conventional pulverized coal-fired boilers, the calcium-based processes' waste product will contain higher levels of unreacted lime; while sodium-based processses will produce wastes with higher levels of soluble sodium compounds. Conventional utility fly ash and scrubber sludge handling procedures including wet sluicing, trucking and ponding, may not be amenable to managing the new waste products. Furthermore, management practices may have a significant impact on process economics, so it is important to develop waste management with disposal and/or utilization practices which minimize handling and disposal costs and maximize the revenue from sales.

The areas which generate the greatest interest for the limestone-based processes include:

- The potential environmental effects associated with handling fine, caustic materials.
- The heat generation caused by the reaction of unreacted lime with atmospheric moisture or process water.
- The corrosive nature of very basic solutions.
- The tendency of fly ash/lime mixtures to set up, or solidify, when exposed to moisture.
- The possibility of basic discharges (pH at or near 12) from storage areas.
- The increased quantities of solid waste.

For the sodium-based processes, the solubility of the sodium compounds is the principal source of waste management concern.

Similar to the SO_2 control processes, coal cleaning generates a solid waste which must be managed in a proper manner. Typically, conventional coal cleaning operations have been situated away from the utility power plant site, with refuse management by others. With the advent of advanced physical coal cleaning technology, it is reasonable to anticipate operations can be located at the power plant site. In such cases, management of the coal cleaning refuse may then become the responsibility of the utility. For coal cleaning refuse, the areas of interest are:

- Acidic nature of leachate.
- Possibility of spontaneous combustion.
- Release of heavy metals in any leachate.

Coal cleaning refuse is significantly different from the conventional coal combustion by-products currently managed by utilities. In fact, it is different from the dry SO_2 technologies discussed above. This difference could have a substantial impact on the cost of managing the refuse at the plant, and to a lesser extent, on the economics of advanced coal cleaning. This impact is important to EPRI because of its research activities at the Coal Cleaning Test Facility, and its impact on utilities which are looking toward the use of advanced physical coal cleaning.

OBJECTIVES OF EPRI RESEARCH PROJECT RP2708

General Program Objectives

In view of the unique characteristics of the solid waste by-products and coal cleaning refuse from the new processes and the potential problems these characteristics may present, EPRI initiated a waste management project in April, 1985, which was preceded by a small planning study. The project, RP2708, is designed to assess the magnitude and potential of waste management problems, formulate preliminary solutions and suggest future EPRI research and development activities to resolve remaining issues, or to verify proposed solutions.

The overall goal of EPRI's Heat Waste and Water Management Program in this area is to develop design and operating guidelines for the management, through disposal and utilization, of solid waste from AFBC and pulverized coal-fired boilers equipped with dry SO_2 controls. EPRI's Coal Quality Program, at its

Coal Cleaning Test Facility (CCTF), has the goal of improving coal cleaning technology, demonstrating new coal cleaning process technologies and characterizing coal cleaning effectiveness on nationally important steam coals.

Specific Objectives of New RP2708 Research Project

The objective of the new RP2708 project is to support the overall goal of both programs by developing technical guidelines which can be used to design and operate a waste management system for each of the five processes described above. In a follow-on study, existing EPRI guideline manuals for ash disposal, sludge disposal, and utilization will be revised, as appropriate, to incorporate the knowledge gained from this research project.

The objective of this research project will be achieved by:

- Gathering and accessing data on solid waste and refuse management systems for fly ash and scrubber sludge from conventional pulverized coal-fired boilers and for conventional coal cleaning operations.
- Identifying the properties of materials that affect the design of each primary piece of equipment.
- Determining those critical properties for the solid wastes and refuse from the dry SO_2 control processes and the advanced physical coal cleaning technologies.
- Developing conceptual designs for solid waste and refuse management systems for each process.
- Identifying further research and development activities necessary to resolve remaining design and operational issues or confirm proposed solutions.

As defined in this presentation and the project, a solid waste management system is all equipment, components, and process streams needed to:

- Remove the waste product from the collection device starting at the hoppers, the thickeners or centrifuge.
- Store or handle the product at the site.

- Transport the product to the ultimate disposal or end-use site.
- Contain the waste at the disposal site, or pretreat the waste before utilization.

The collection device itself, either the electrostatic precipitator, fabric filter, thickener or centrifuge is not included in this definition, or in the research project.

The project objective will be met by completing the five major tasks discussed briefly in the following sections.

Background Information Survey

This activity is a survey, compilation and evaluation of information on conventional solid waste and coal cleaning refuse properties; management, disposal and utilization practices; and, on the properties of the solid waste products and refuse from the new processes. The survey is designed to gather information on:

- Conventional fly ash, bottom ash, scrubber sludge and coal cleaning refuse handling systems, storage systems, transportation systems, disposal sites, and on the ash and refuse properties affecting design.
- Current Federal and state regulations affecting design and operation of the disposal site or utilization practice, along with any uncertainties in the regulatory area.
- Compositional and behavioral aspects of both the solid wastes from the new SO_2 control processes and the advanced physical coal cleaning technologies.

The survey is designed to account for the variation in information available from one technology to another. For example, limited survey effort is required for the AFBC option because of the Alliance 6' x 6' facility and the 20 MWe facility at TVA's Shawnee steam plant. For the advanced physical coal cleaning technology, the survey will use data from conventional coal cleaning, as no advanced systems currently are in operation.

An important work activity during this survey will be to identify preferred operating conditions necessary to obtain data during the solid waste characterization work. A complete set of preferred operating conditions will be offered to facilities which are going to be the source of samples for characterization work. The operating conditions will reflect stage of development, flexibility of the facilities' operating range, and minimum sample variability, among other factors.

Solid Waste Characterization

The solid waste characterization task is a detailed characterization of solid waste products and refuse based on an evaluation of data gathered above, and analysis of samples gathered at appropriate test, pilot and full-scale facilities. Usefulness and limitations of standard laboratory property determinations as they apply to the dry SO_2 processes and the advanced physical coal cleaning technologies will also be investigated.

The characterization will provide compositional and behavioral properties relating to the design and operation of a solid waste management system, as well as properties relating to potential utilization options. Data voids identified in the information survey will be filled, existing data augmented and where necessary, updated to reflect advances in a particular SO_2 control process. Specifically, the characterization will provide data necessary to:

- Predict solid waste behavior during handling, storage, transport and disposal.
- Establish general design and operational guidelines for the solid waste management system of each dry SO_2 control process, and the coal cleaning technology.
- Estimate the quantity and quality of both leachate and surface water runoff.
- Evaluate potential utilization options.

It is intended that each SO_2 control process and the coal cleaning process will be sampled at least five times during the project. An exception to this case will be for spray drying, which will have eight events to account for once-through and recycle systems, as well as high and low-alkalinity ash.

This characterization task is extremely critical to the success of the research project, as it impacts each of the three remaining tasks. To be as comprehensive and responsive as practicable, the work activity is divided into four subject areas:

- A detailed sampling and analysis plan.
- Analysis of the chemical composition of materials.
- Determining the physical and engineering properties.
- Estimating the leachate and runoff characteristics.

Sampling and Analysis Plan. Requirements for the sampling and analysis plan are extensive, and a separate plan is expected for each of the five technologies in the project. At a minimum, each plan will address scheduling as it relates to activities at the sample site, sampling procedures, quantity of sample required, field and laboratory quality control and assurance, data management and analytical techniques to be used. During the development of the plan and during the characterization work, the project will have test procedures evaluated to determine their applicability to samples with high calcium or sodium content, and to see that they provide adequate information for the waste management system design.

Chemical Composition of Materials. Table 1 is a listing of the elemental chemical analyses to be performed on representative solid waste samples and the corresponding coal ash. While there is no regulatory or industry standard set of elemental and anionic species analysis, those in Table 1 are considered significant in selecting materials of construction, estimating ranges for engineering properties and assessing compliance with surface and groundwater limitations. For the calcium based processes, the corrosive nature,

Table 1

LIST OF CHEMICAL SPECIES INCLUDED IN ANALYSIS[a]

Parameter

Aluminum
Antimony
Arsenic
Barium
Beryllium
Boron
Cadmium
Calcium
Chromium
Cobalt
Copper
Fluoride
Iron
Lead
Lithium
Manganese
Magnesium
Mercury
Molybdenum
Nickel
Potassium
Selenium
Silver
Sodium
Strontium
Thallium
Tin
Titanium
Uranium
Vanadium
Silica
Zinc
Hydroxide
Carbonate
Total Soluble Sulfur
Sulfate
Sulfite
Total Soluble Chloride
*pH
*Chloride
*Nitrate
*Nitrite
*Total Dissolved Solids
*Cyanide
*Bromine
*Phosphorus
*Total Carbon
*Total Inorganic Carbon
*Total Organic Carbon

Note: [a] All parameters would be included in the leachate tests. Those with asterisks would not be analyzed under chemical composition of materials.

reactivity and tendency to solidify with increasing moisture content will also be determined. The coal cleaning refuse will be subjected to the analyses in Table 2 as an indicator of recoverable coal.

For the solid wastes, refuse and coal ashes, activities of gross alpha, gross beta, Ra^{226}, and U^{235} will be measured. Also, an analysis program has been established to determine whether certain classes of organic compounds are present on coal ashes, ash from coal cleaning refuse

Table 2

ANALYSIS OF COAL SAMPLES[a]

Analysis	Result
Proximate Analysis	Ash, % by weight Volatile Matter, % by weight Fixed Carbon, % by weight
Heating Value	Btu per pound
Ultimate Analysis	Carbon, % by weight Hydrogen, % by weight Nitrogen, % by weight Sulfur, % by weight Ash, % by weight Chlorine, % by weight

---

Note: [a] Analysis to be on a dry basis.

and the solid wastes from the dry SO_2 control processes. The analysis will be for both the ash and refuse as received and for the leachate from extraction and column tests. The analysis will be for those compounds listed in Table 3. These compounds are part of the U.S. EPA's priority pollutants list and have been reported as being present on some of the materials being studied in this project. It should be noted that this list serves as a point of reference only, and like the radioactivity of coal combustion wastes the information is being obtained to establish a data base should any regulatory action be forthcoming.

Physical and Engineering Properties. The physical and engineering properties of the solid waste materials will provide data to develop design and operating guidelines for the waste management systems, and for some potential utilization options. These same properties also can be used to

Table 3

ORGANIC COMPOUNDS FOR ANALYSIS

Compound	Source	Compound	Source
Acenaphthene	C	Chrysene	F,B,C
2,4-dimethylphenol	F	Acenaphythylele	B,C
Fluoranthene	F,B,C	1,12-benzoperylene (benzo(ghi)berrylene)	B,C
Naphthalene	F,B,C	Fluorene	B,C
Diethyl phthalate	W	Phenanthrene	F,B,C
Dimethyl phthalate	W	1,2,5,6-dibenzanthracene (dibenzo(h)anthracene)	B,C
Benzo(a)pyrene (3,4-benzo-pyrene)	B,C	Pyrene	F,B,C
3,4-benzofluoanthene (benzo(b)fluoranthene)	C		

Source: Envirosphere Company, 1985.

Legend: B - Bottom Ash, C - Coal Ash, F - Fly Ash, W - Wet Ash

help predict the formation and transport of leachates. Table 4 is a list of the physical and engineering properties to be determined. A detailed approach will be developed to reconcile the need to conduct engineering property tests at high moisture contents when the samples will most likely be in a very dry state.

Leachate and Runoff Characteristics. The leachate and runoff from a dispoasl area are significant waste management considerations. The significance results from the potential for release of chemical species into surface or groundwaters at concentration levels at or near regulatory limits. Prediction of leachate and runoff quality from conventional coal combustion waste sites, as well as coal cleaning refuse, is still tenuous. However, it is necessary to be able to apply scientific and engineering judgment to selected test results and site conditions to assess potential impacts and, therefore, to

Table 4

LIST OF PHYSICAL AND ENGINEERING PROPERTIES TO BE DETERMINED

Property	Unit of Measure
Crystal Morphology	--
Bulk Density	lbs/ft^3
Particle Size Distribution	percent finer than stated size
Specific Gravity	dimensionless
Moisture Content	percent by weight
Permeability	cm/sec or ft/yr
Specific Surface Area	area per mass, m^2/g
Angle of Interal Friction	degrees
Dry Unit Weight	lbs/ft^3
Atterberg Limits	--
Compaction	--
Compressibility and Consolidation	--
Unconfined Compressive Strength	--
Triaxial Compression Test	--

provide mitigative measures. Accordingly, the leachate testing, as presently planned, will include the standard static extraction test and a dynamic test using long-term column tests. In both cases, the chemical analyses to be included are listed in Table 1. In addition, the proposed Toxicity Characteristic Leaching Procedure (TCLP) unveiled by the U. S. Environmental Protection Agency in April, 1985 will be tested on the advanced SO_2 wastes.

Solid Waste Management System Conceptual Designs

This work activity will develop conceptual solid waste management system designs for each of the dry SO_2 processes, and for a wet and dry advanced physical coal cleaning process. The impact of unique properties of waste products on the design, operation, and costs of waste management systems will be estimated and compared to waste management systems for conventional pulverized coal-fired plants and conventional coal cleaning facilities. This task includes conceptual designs for new plants, a retrofit and a preliminary liner evaluation.

The conceptual designs will include general site plans, flow diagrams of the solid waste management system and general arrangement drawings for the waste handling systems, transportation systems (if appropriate), storage facilities (if needed), and disposal site. Drawings will contain sufficient detail to confirm engineering assumptions and to support cost estimating. Estimated costs will be developed to show total capital cost, total annual operating and maintenance cost, and levelized annual cost for the life of the plant. An estimated incremental levelized annual cost over that of a conventional management system will be developed.

To demonstrate economic impact, a levelized cost (including operating, maintenance and capital) per kWh for the solid waste management system and an incremental cost per kWh over a conventional system will be developed. Finally, a total cost per ton of solid waste generated will be calculated. This total cost will also be levelized over the life of the plant.

In developing conceptual designs, the following will be considered:

- In-plant waste handling including dry versus wet, segregation of different waste types, potential for reactive fugitive dust, and personnel health and safety.
- Temporary storage including size, type of vessel or facility, moisture content, heat of hydration, and transfer system.
- Transportation to disposal site including wet versus dry transport, reactive fugitive dust, waste hauling on public roads, and weather conditions.
- Disposal including ponds versus landfills, requirements for pretreatment, mechanical properties, water quality criteria, type of base (or liner), and method of waste placement.

By-Product Utilization

Activities for by-product utilization will determine if the unique properties of the solids from dry SO_2 control processes will affect their suitability for conventional uses of fly ash or scrubber sludge, and identify potential new uses which take advantage of these properties. The potential uses for coal cleaning refuse from the advanced technologies will also be examined and compared to current practices.

An assessment framework will be developed to provide a rationale for identifying potential end-uses of these new materials. As a beginning point, the current uses of conventional ash and scrubber sludge will be evaluated relative to the new wastes to be generated. Examples of potential applications include structural fill, pozzolanic base course, mineral filler in asphaltic mixes, soil stabilization and/or amendment, grout, cement manufacturing, concrete products and AFBC boiler fuel (coal cleaning waste), among others. Other uses for ash are in various stages of development and will also be used; examples are alumina extraction, plastics extender and metals recovery.

For each potential utilization, an analysis of research findings will be made in terms of the critical factors influencing the suitability of the given by-product. Based on this analysis and the similarity or conformance with accepted standards and guidelines, justification will be made to either abandon the utilization on technical grounds, or recommend further development effort. For those meriting further development, the by-product will be evaluated to determine:

- Range of properties acceptable for the particular end-use.
- Approximate size and location of market.
- Relative value of the by-product compared to the virgin material.
- Transportation requirements.
- Special processing equipment.

Case Studies

Case studies will develop preliminary solid waste system designs for three specific processes and plant sites to support other EPRI research and development studies, and to demonstrate and refine the conclusions from the conceptual studies and the utilization analysis. It is anticipated the case studies will include a 50 MWe pulverized coal-fired boiler with limestone injection, a 500 MWe pulverized coal-fired boiler with sodium injection, using a baghouse for particulate control, and a 20 MWe AFBC facility.

The case studies will contain similar information to that in the conceptual design, yet with more site specific detail. Also, the case studies will consider the changes that must be made to the existing system in order to deal with the new waste material. Alternate subsystems and operational procedures will be specified in sufficient detail to allow cost estimates to be made. As an alternate system, the site will remain the same but an assumption will be made that the ash handling system is to be completely replaced.

SELECTED RESULTS TO DATE

To initiate RP2708 in a timely manner, a preliminary information survey and data evaluation project was completed. The objectives of the preliminary project were to:

- Compile information on the design and operation of conventional solid waste management practices and on the properties of the waste disposal products from the new SO_2 control processes.
- Examine the information obtained above to identify the properties that affect the design and operation of a waste management system.
- Evaluate the available property data for the dry SO_2 control processes to assess its usefulness.
- Identify those chemical, physical and engineering properties of the wastes from the new dry SO_2 control processes that are not well known but are needed to design and operate a solid waste management system.

The remaining sections of this presentation provide a brief summary of the results from the preliminary study, along with a discussion of the approach used to develop the results.

APPROACH

In a step-wise fashion, the approach to identifying those chemical, physical and engineering properties of the wastes from the new dry SO_2 control processes which are not well know but are needed to design and operate a solid waste management system was as follows:

- Step 1 was to establish a logical breakdown of the conventional solid waste management system into subsystems.
- Step 2 was to identify the design criteria and operating guidelines for conventional systems which are associated with the subsystems, along with the applicable test or analytical procedure.
- Step 3 was to develop a list of all chemical, physical and engineering properties for conventional wastes which have been measured and for which data are reported in the literature.
- Step 4 was to match the list from Step 3 with the design criteria and guidelines for Step 2; thereby, identifying those conventional waste characteristics significant for design and operation.
- Step 5 identifies those chemical, physical and engineering properties from the new SO_2 control technologies for which there is no reported data or only limited data.

Review of Conventional Systems

As the first step in arriving at those parameters which could be significant in the design of a waste management system for dry SO_2 control technologies, key design criteria for a conventional waste system were identified. The conventional waste management system for a pulverized coal-fired unit was considered to include all equipment and components needed to: (a) remove the waste product from the collection device starting at the hoppers; (b) store or handle the product at the site; (c) transport the product to the ultimate end-use site; and, (d) contain the waste at the disposal site or pretreat the waste before utilization. The collection device itself was not included as part of the system.

The various subsystems involved in fly ash management were isolated and were identified as handling, storage, transport, disposal and utilization. The information survey for the conventional system focused on fly ash, and the design criteria for each subsystem were identified. The design criteria were determined from literature review, discussions with vendors and manufacturers, EPRI design manuals, relevant experiences from the lime, cement and soda ash industry, and engineering experience. The design criteria, once determined, were then associated with a chemical characteristic, a physical characteristic or an engineering property.

Table 5 is a subsystem breakdown for the conventional waste management system, along with the general design criteria associated with each subsystem. Because of the similarities in design criteria for handling, storage and transport, these subsystems were placed into one category. Disposal and utilization were addressed as separate entities. The criteria are delineated by chemical, and physical and engineering groupings for each subsystem category.

An examination of Table 5 indicates there are not an abundance of chemical, and physical and engineering criteria for the handling, storage and transport subsystems. Disposal and utilization, on the other hand, have a much larger set of criteria used for design purposes. A major factor influencing this larger set for disposal is the need to afford adequate protection for surface and ground water, within various regulatory frameworks. Utilization has a large set of criteria because of the relatively large number of different end-uses, standards and guidelines for those end-uses, the need to ensure quality control of a by-product to minimize liability, and the institutional constraints of a by-product having to compete with an established virgin material, often represented by a strong trade association.

Table 5

DESIGN CRITERIA FOR CONVENTIONAL HANDLING, TRANSPORT AND STORAGE SUBSYSTEMS

Criteria	Parameter	Associated Test or Standard (1)
Corrosivity	pH	Chemical Analysis
	Corrosivity	CEMA B-4
Abrasivity	Abrasion	CEMA A-16, ASTM D658
	SiO_2	ASTM D2795-69
	Al_2O_3	ASTM D2795-69
	Fe_2O_3	ASTM D2795-69
	Particle Shape	CEMA B-17
Cementitiousness	CaO	ASTM C114-69
Unburned Carbon	LOI	ASTM C311-68
Mininimum Velocity for Pneumatic Transport	Particle Size	ASTM D854-58
	Temperature	In-Situ Measurement
Fluidizability	Particle Size Distribution	ASTM D422-73
	Granular vs. Powder	Calculate from Size Distribution
	Temperature	In-Situ Measurement
Bulk Density	Moisture Content	ASTM D2216-71
	Particle Size Distribution	ASTM D422-73
	Specific Gravity	ASTM D854-58
	Angle of Repose	ASTM D3080-72
Fugitive Dust	Particle Size Distribution	ASTM D422-73
	Granular vs. Powder	Calculate from Size Distribution
	Specific Gravity	ASTM D854-58
Working Volumes	Aerated Bulk Density	CEMA A-8
Stored Volumes	Packed Bulk Density	CEMA A-9
Structural Design	Packed Bulk Density	CEMA A-9

Table 5 (Continued)

DESIGN CRITERIA FOR CONVENTIONAL HANDLING, TRANSPORT AND STORAGE SUBSYSTEM

Criteria	Parameter	Associated Test or Standard (1)
Slope of Hoppers and Bins	Angle of Repose	ASTM D3080-72
Angle of Surcharge on Conveyors	Angle of Repose	ASTM D3080-72
Side Slopes of Piles	Angle of Repose	ASTM D3080-72
Self-Hardening	CaO	Chemical Analysis
	Temperature	In-Situ Measurement
	Strength over Time	ASTM D2166-66, ASTM D2435-70, ASTM D3080-72

DESIGN CRITERIA FOR CONVENTIONAL DISPOSAL SUBSYSTEM

Criteria	Parameter	Associated Test or Standard (1)
Long-Term Strength	Na_2O	Chemical Analysis
	K_2O	Chemical Analysis
	Ca as Available Ca	Chemical Analysis
	MgO	Chemical Analysis
	LOI	ASTM C311-68
	Particle Size Distribution	ASTM D422-73
	Aerated Bulk Density	CEMA A-8
	Packed Bulk Density	CEMA A-9
	Angle of Repose	ASTM D3080-72
Water Quality Standards	pH	Chemical Analysis
	Al, Sb, As, Ba, Be, B	Chemical Analysis
	Br, Cd, Cl, Cr, Co, Cu	Chemical Analysis
	F, I, Fe, Pb, Li, Mn	Chemical Analysis
	Hg, Mo, Ni, NO_3,	Chemical Analysis
	NO_2, Se, Ag, Sr, SO_4	Chemical Analysis
	Th, Sn, U, Va, Zn	Chemical Analysis

Table 5 (Continued)

DESIGN CRITERIA FOR CONVENTIONAL DISPOSAL SUBSYSTEM

Criteria	Parameter	Associated Test or Standard (1)
Permeability	Permeability	ASTM D2434-68
	Particle Size Distribution	ASTM D422-73
	CaO	Chemical Analysis
	Aerated Bulk Density	CEMA A-8
	Packed Bulk Density	CEMA A-9
Settling Characteristics of Suspended Solids in Runoff	Particle Size	ASTM D854-58
	Specific Gravity	ASTM D854-58
Abrasion of Liners	Sphericity	CEMA B-17
	Abrasivity	CEMA A-16
Ion Exchange and Adsorption	Specific Surface Area	BET Nitrogen Adsorption Method
Density	Moisture Content (Natural and Optimum)	ASTM D2216-71, ASTM D690-70, ASTM D1557-70
Steepness of Slopes	Shear Strength	ASTM D2166-66, ASTM D3080-72, ASTM D2435-70
	Angle of Repose	ASTM D3080-72
Leachate Production	pH	Chemical Analysis
	Leachable Alkalinity as CaO	Chemical Analysis
	Fe_2O_3	Chemical Analysis
	Capillary Rise	No Standard Reported
	Water-Soluble Content	Chemical Analysis
	Leachate Formation	EPA Extraction Procedure

Table 5 (Continued)

DESIGN CRITERIA FOR CONVENTIONAL
UTILIZATION SUBSYSTEM

Criteria	Parameter	Associated Test or Standard (1)
Strength	Na_2O	Chemical Analysis
	K_2O	Chemical Analysis
	SiO_2	Chemical Analysis
	Al_2O_3	Chemical Analysis
	Fe_2O_3	Chemical Analysis
	CaO	Chemical Analysis
	MgO	Chemical Analysis
	LOI	ASTM C311-68
	Aerated Bulk Density	CEMA A-8
	Packed Bulk Density	CEMA A-9
	Shear Strength	ASTM D2166-66, ASTM D3080-72, ASTM D2435-70
	Capillary Rise	No Standard Reported
	Unconfined Comp. Strength	ASTM D2166-66
Self-Hardening	Na_2O	Chemical Anslysis
	K_2O, CaO	Chemical Analysis
ASTM Limits for Use In Concrete	SO_3	ASTM C618-78
	SiO_2	ASTM C618-78
	Al_2O_3	ASTM C618-78
	Fe_2O_3	ASTM C618-78
	CaO	ASTM C618-78
	MgO	ASTM C618-78
	Cesium, Li, Ru	ASTM C618-78
	Moisture Content-Natural	ASTM C618-78
	LOI	ASTM C618-78
Soil Amendment Properties	pH	Chemical Analysis
	Porosity	Calculated from bulk density and specific gravity
Leachate Creation	Leachable Alkalinity as CaO	Chemical Analysis

Table 5 (Continued)

DESIGN CRITERIA FOR CONVENTIONAL UTILIZATION SUBSYSTEM

Criteria	Parameter	Associated Test or Standard (1)
Ingredient for Fertilizer	CaO	Chemical Analysis
	MgO	Chemical Analysis
	P_2O_5, B, Se	Chemical Analysis
	Cu, Zn, Fe, Mn	Chemical Analysis
	SO_4	Chemical Analysis
	NO_3	Chemical Analysis
Dewatering Potential for Sewage Sludge	Particle Size	ASTM D854-58
	Particle Size Distribution	ASTM D422-73
Filter Compatibility with Underdrains as Fill	Particle Size	ASTM D854-58
	Permeability	ASTM D2434-68
Use as Cement or Asphalt Replacement	Particle Size Distribution	ASTM D422-73
	Pozzolanic Potential	ASTM C311-77, ASTM 595-79
	Self-Hardening Potential	Shear Strength Over Time
Use as Grout	Durability-Uniformity Coefficient	Calculate from size distribution
	Flow Properties-Sphericity	CEMA B-17
Settlement of Fills	Particle Specific Gravity	ASTM D854-58
	Compressibility	ASTM D2435-70
Steepness of Slopes	Shear Strength	ASTM D3080-72

Table 5 (Continued)

DESIGN CRITERIA FOR CONVENTIONAL
UTILIZATION SUBSYSTEM

Criteria	Parameter	Associated Test or Standard (1)
Fill Bearing Capacity	Shear Strength	ASTM D2166-66, ASTM D3080-72, ASTM D2435-70
Fill Stability	Capillary Rise	No Standard Reported
	counters)	

Source: Envirosphere Company, 1985.

(1) ASTM - American Society for Testing and Materials
CEMA - Conveyor Equipment Manufacturer's Association

Another factor which appears to limit the number of significant design criteria for handling, storage and transport is the experience factor of the utility industry, engineer and manufacturer/vendor. Much of the materials of construction and the system components have been subject to a number of years of on-line use. Operator and manufacturer/vendor experience has been able to account for the typical changes in waste characteristics, and only a few characteristics remain as criteria requiring actual testing or analysis.

Once the criteria and guidelines were identified, the next step was to determine the expected range for a particular parameter, make sure of the reason for its significance and to obtain a reasonably large set of data to establish the range. Where appropriate, the quality of the data was generally assessed by size of sample, source of sample and nature of reporting. A matrix was developed grouping the various subsystems with the significant design criteria. Within the matrix, the headings indicated parameter or property of interest, range of the data, reason or significance of the parameter or property, any data qualifiers and, finally, the reference for the data. A series of tables were prepared which pertained to all the technologies, the individual subsytems, and parameters. An example is presented in Table 6 for handling, storage and transport subsystems for the conventional pulverized coal-fired boiler.

Review of Each of the New Technologies

The matrices from the conventional waste management systems were used as a basis for gathering and evaluating data for the dry SO_2 control technologies. Information concerning each of the four processes, i.e., spray drying, AFBC, limestone furnace injection and dry sodium injection, were reviewed with the intent being to obtain characteristics matching those in the matrices for the conventional waste management systems. These characteristics were for both the chemical parameters, and the physical and engineering properties. Information for the new technologies, including data range, any data qualifiers and the references for the data, were superimposed upon the conventional waste management matrix. A number of tables were prepared to illustrate this matrix. Table 7 is an example of one of these tables for the the AFBC technology. Because the objective was to cover the full range of values the waste material could exhibit, the ranges shown for the new technology are representative of all waste materials that individual process produces. For example, the ranges shown on Table 7 (the upper range is for the conventional waste and the lower range for the new process) include the characteristics of both spent bed material and fly ash produced by the AFBC process.

A review of the matrix developed for each new technology, by subsystem and parameter, permitted the identification of how the waste characteristics compared with the related conventional waste ranges. Table 8 shows the results of this comparison for the AFBC technology. The various subsytems are

Table 6

HANDLING, TRANSPORT, AND STORAGE SUBSYSTEM CONVENTIONAL MATERIAL - FLY ASH

Chemical Parameter	Data Range	Reason or Significance	Data Qualifiers and References
pH	3-12	Corrosivity	Data from Ref.5 based on a set of 3 samples.
SiO_2	6-68%	Abrasivity	Ref.1 is from experience and literature search, Ref.8 had large number of samples, while Ref.2 is for 14 samples.
Al_2O_3	4-44%	Abrasivity	Source of coal for Ref.7 not identified, the range was anthracite to lignite, Ref.8 had large number of samples.
Fe_2O_3	1-44%	Abrasivity	Ref.2 is for 14 samples, see notes above for Refs.7 and 8.
CaO	0.2-52%	Cementitious	Source of coal for Ref.7 not identified, the range was anthracite to lignite, Ref.14 had large number of samples.
Loss on Ignition	0.32-20.5%	Indicates Unburned Carbon	Ref.2 is for 14 samples.

Table 6 (Continued)

HANDLING, TRANSPORT, AND STORAGE SUBSYSTEM CONVENTIONAL MATERIAL - FLY ASH

Physical/ Engineering Parameter	Data Range	Reason or Significance	Data Qualifiers and References
Particle Size	1-400 micron	Determines minimum velocity for pneumatic transport	Ref.1 is from experience and literature search, Ref.4 is a literature search.
Particle Size Distribution	See Fig. 1	Fluidizability, fugitive dust, affects bulk density	Ref.1 is from experience and literature search.
Granular vs. Powder (Divide at 200 mesh = 0.074 mm)	No Data	If more than 30% powder it can be fluidized, fugitive dust	Ref.1 is from experience and literature search.
Particle Specific Gravity	1.5-3.1	Fugitive dust, affects bulk density	Ref.1 is from experience and literature Ref.3 had large number of samples.
Bulk Density (Aerated)	35-65 pcf	To calculate working volumes, bin and silo sizing	Ref.1 is from experience and literature search.
Bulk Density (Packed)	60-105 pcf	To calculate stored volumes, structural design	Ref.1 is from experience and literature search, Ref.8 had large number of samples.
Angle of Repose	35-55 degrees	Slope of hoppers and bins for gravity flow, angle of surcharge on conveyors, affects bulk density, side slopes of piles	Ref.1 is from experience and literature search.
Moisture Content (Natural)	0.06-5%	Density, affects self-hardening	Ref.2 is for 14 samples, Ref.4 is a literature review.

Table 6 (Continued)

HANDLING, TRANSPORT, AND STORAGE SUBSYSTEM
CONVENTIONAL MATERIAL - FLY ASH

Physical/ Engineering Parameter	Data Range	Reason or Significance	Data Qualifiers and References
Abrasion	No Data	Wears out equipment	
Temperature	No Data	High temperature lowers air viscosity and reduces carrying capacity, low temp. increases se[illegible]hardening and can cause freezing	
Fluidizability	No Data	Fluidizable, if more than 30% powder, conical vs flat-bottomed hoppers	
Self-Hardening	0 day 257 - 2134 psf, 7 day 316 - 1994 psf, 28 day 256 - 3228 psf	Clogging	Ref.6 is for a set of 12 samples.

Source: Envirosphere Company, 1985.

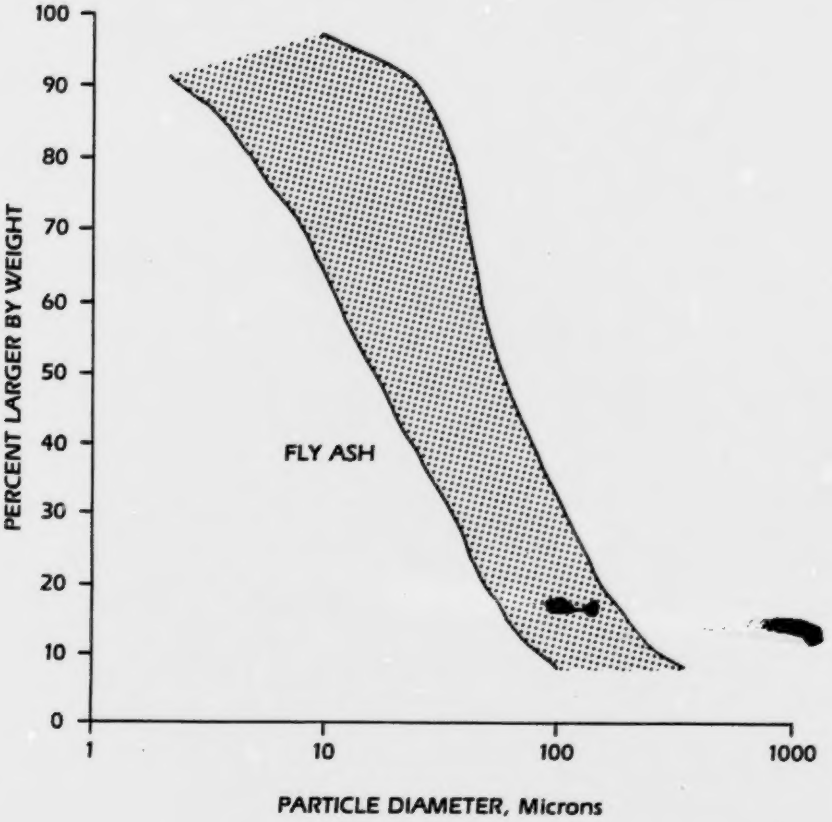


FIGURE 1

GRAIN SIZE DISTRIBUTION CURVE CONVENTIONAL FLY ASH

Table 7

HANDLING, TRANSPORT, AND STORAGE SUBSYSTEM
CONVENTIONAL MATERIAL - FLY ASH (above line)
NEW TECHNOLOGY MATERIAL - AFBC (below line)

Chemical Parameter	Data Range	Reason or Significance	Data Qualifiers and References
pH	3-12	Corrosivity	Ref.6 is for a set 3 samples.
	12.3-12.5		Ref.13 from 2 pilot studies.
SiO_2	6-68%	Abrasivity	Ref.1 is from experience and literature search, Ref.8 had large number of samples, Ref.2 is for 14 samples.
	22.4-31.2%		Ref.11 from two pilot plants.
Al_2O_3	4-44%	Abrasivity	Source of coal for Ref.7 not identified, the range was anthracite to lignite, Ref. 8 had large number of samples.
	5.9-8.4%		Ref.11 from two pilot plants.
Fe_2O_3	1-44%	Abrasivity	Ref.2 is for 14 samples, see notes above for Refs.7 and 8.
	2.1-11%		Ref.9 for pilot study, Ref.11 from two pilot studies.
CaO	0.2-52%	Cementitious	See notes above for Refs.7 and 8.
	3.9-56.3%		Ref.9 for pilot study, Ref.11 from two pilot studies.
Loss on Ignition	0.32-20.5%	Indicates Unburned Carbon	Ref.2 is for 14 samples.
	0.3-2.1		Ref.10 from pilot study.

Table 7 (Continued)

HANDLING, TRANSPORT, AND STORAGE SUBSYSTEM
CONVENTIONAL MATERIAL - FLY ASH (above line)
NEW TECHNOLOGY MATERIAL - AFBC (below line)

Physical/ Engineering Parameter	Data Range	Reason or Significance	Data Qualifiers and References
Particle Size	1-400 micron 250-850 microns	Determines minimum velocity for pneumatic transport	Ref.1 is from experience and literature search, Ref.4 is a literature search. Ref.14 is a literature search.
Particle Size Distribution	-	Fluidizability, fugitive dust, affects bulk density	Ref.1 is from experience and literature search.
Granular vs. Powder (Divide at 200 mesh = 0.074 mm)	No Data	If more than 30% powder it can be fluidized, fugitive dust	Ref.1 is from experience and literature search.
Particle Specific Gravity	1.5-3.1 1.94-2.63	Fugitive dust, affects bulk density	Ref.1 is from experience and literature search, Ref.3 had large number of samples. Ref.9 is a pilot study.
Bulk Density (Aerated)	35-65 pcf 14-91 pcf	To calculate working volumes, bin and silo sizing	Ref.1 is from experience and literature search. Ref.9 is a pilot study.
Bulk Density (Packed)	60-105 pcf 37-100 pcf	To calculate stored volumes, structural design	Ref. 1 is from experience and literature search, Ref.8 had large number of samples. Ref.9 is a pilot study.

Table 7 (Continued)

HANDLING, TRANSPORT, AND STORAGE SUBSYSTEM
CONVENTIONAL MATERIAL - FLY ASH (above line)
NEW TECHNOLOGY MATERIAL - AFBC (below line)

Physical/ Engineering Parameter	Data Range	Reason or Significance	Data Qualifiers and References
Angle of Repose	35-55 degrees	Slope of hoppers and bins for gravity flow, angle of surcharge on conveyors, affects bulk density, side slopes of piles	Ref.1 is from experience and literature search.
Moisture Content (Natural)	.06-5% 0%	Density, affects self-hardening	Ref.2 is from set of 14 samples, Ref.4 is a literature search. Ref.12 is a literature search and four pilot plants.
Abrasion	No Data	Wears out equipment	
Temperature	No Data	High temperature lowers air viscosity and reduces carrying capacity, low temp. increases self-hardening and can cause freezing	
Fluidizability	No Data	Fluidizable, if more than 30% powder, conical vs flat-bottomed hoppers	
Self-Hardening	0 day 257-2134 psf, 7 day 316-1994 psf, 28 day 256-3228 psf	Clogging	Ref.6 is for a set of 12 samples.

Source: Envirosphere Company, 1985.

Table 8

ATMOSPHERIC FLUIDIZED BED COMBUSTION DATA MATRIX

Physical/Engineering Parameters	Handling, Storage and Transport	Disposal	Utilization
Particle Size	A	A	A
Specific Gravity	W	W	W
Packed Bulk Density	B	B	B
Moisture Content*	W	W	W
Unconfined Compressive Strength	-	-	NR
Permeability	-	A	A
Radioactivity	-	NR	NR
Aerated Bulk Density	A,B	A,B	A,B
Angle of Repose	NR	NR	-
Self-Hardening	NR	NR	NR
Specific Surface Area	-	B	-
Shear Strength	-	NR	NR
Capillary Rise	-	NR	NR
Water Soluble Content	-	NR	-
Pozzolanic Potential	-	-	NR
Compressibility	-	-	NR
Optimum Moisture Content	-	NR	NR

*Moisture content at the point of collection

Chemical Parameter			
P_2O_5	-	-	NR
SiO_2	W	-	W
Al_2O_3	W	-	W
Fe_2O_3	W	W	W
CaO	A	A	A
LOI	W	W	W
pH	A	A	A
Na_2O	-	NR	NR
K_2O	-	W	W
MgO	-	W	W
Aluminum	-	B	-
Antimony	-	B	-
Arsenic	-	B	-
Barium	-	B	-
Beryllium	-	B	-
Boron	-	B	-

Table 8 (Continued)

ATMOSPHERIC FLUIDIZED BED COMBUSTION DATA MATRIX

Chemical Parameters	Handling, Storage and Transport	Disposal	Utilization
Bromide	-	A	-
Cadmium	-	W	-
Cesium	-	-	B
Chloride	-	A,B	-
Chromium	-	W	-
Cobalt	-	B	-
Copper	-	B	B
Fluoride	-	W	-
Iodine	-	W	-
Iron	-	B	B
Leachable Alkalinity	-	NR	NR
Lead	-	B	-
Lithium	-	B	W
Manganese	-	W	B
Mercury	-	A	-
Molybdenum	-	B	-
Nickel	-	A	-
Nitrate	-	W	W
Rubidium	-	-	B
Selenium	-	B	-
Silver	-	W	-
Strontium	-	W	-
Sulfate	-	A	A
Thallium	-	W	-
Tin	-	W	-
Uranium	-	W	-
Vanadium	-	W	-
Zinc	-	B	B

Symbols:
A: Can be above conventional range.
B: Can be below conventional range.
W: Within conventional range.
NR: Not reported.
- : Not applicable under particular subsystem.

addressed by chemical, and physical and engineering parameters. The reported data was identified as being within the range of conventional waste, above or below the range, not applicable to the subsystem, or having no reported data. As seen in Table 8, several physical and engineering properties of AFBC wastes need to be determined. It is significant to note that the AFBC technology represents a new technology which has received as much attention, if not more than any of the others.

SUMMARY

Although a complete description of the results of the comparison of characteristics between new technology waste materials and conventional technology waste materials is beyond the scope of this paper, some general conclusions can be drawn. As would be expected, the data on properties of waste materials from conventional technology are more extensive than those from the new processes, both with respect to the number of characteristics reported, and with respect to the reliability present in the number of observations for each characteristic. The conventional systems data are also more reliable because they were obtained from a higher proportion of prototypes, as opposed to the data for new process materials obtained largely from pilot studies. For these reasons, the fact a new process parameter has been reported within the range of conventional material cannot be assumed to indicate it will always fall within that range.

The data indicate new process waste material is generally composed of two fractions, one of which is derived from coal combustion and is chemically similar to conventional fly ash, and the other which is derived from material added within the new process. The latter fraction includes major chemical components of the additive and reaction products of sulfur removal, as well as trace materials derived from impurities in the additive. For example, in the case of AFBC, the chemical parameters whose ranges exceed those reported for conventional fly ash include pH, calcium, bromide, chloride, mercury, nickel and sulfate. The pH, calcium and sulfate are assumed to be direct results of the process, while the other parameters could be the result of impurities in the bed material, i.e. limestone, or the lower burner temperatures, or the coal source.

With respect to physical characteristics, in particular within the handling, storage, and transport subsystems, it appears there are few data available even for conventional systems. These systems are generally designed by vendors, based on previous experience. If a parameter causes a problem by exceeding the vendors factors of safety, case-by-case solutions are found. Finally, the data on new processes focus mainly on demonstrating the process does remove sulfur, rather than on waste management design and operating considerations.

For the above reasons, it is concluded that extensive testing of new process waste materials should continue, especially as larger scale projects are constructed and more representative materials are produced. Such a broadening of the new process data base, both with respect to the number of parameters reported and to the number of samples tested, would help to establish a reasonable set of criteria for use in preliminary design, case studies and examining utilization options.

REFERENCES

1. Ash Systems Engineering, Inc. Ash Handling System Reference Manual, RP 1835-4, Final Report (Draft). Prepared for Electric Power Research Institute.

2. Radian Corporation. Chemical/Physical Stability of Flue Gas Cleaning Wastes, EPRI FP-671, Project 786-2, Final Report, Volume 2 January, 1979. Prepared for Electric Power Research Institute.

3. Tetra Tech, Inc. Physical-Chemical Characteristics of Utility Solid Wastes, EPRI EA-3236, Project 1487-12, Final Report, September, 1983, prepared for Electric Power Research Institute.

4. Michael Baker, Jr. Inc., Stabilization of Power Plant By-Product Storage Sites for Building Development, EPRI CS-3475, Project 1685-6, Final Report, April, 1984. Prepared for Electric Power Research Institute.

5. Matrecon, Inc. Leachate Control and Monitoring Systems for Solid Waste, RP 1457-1, Interim Report, January, 1983. Prepared for Electric Power Research Institute.

6. Tennessee Valley Authority. Economic Analysis of Wet Versus Dry Ash Disposal Systems, Interagency Energy/Environment R&D Program Report, EPA-600/7-81-013, January, 1981. Prepared for U.S. Environmental Protection Agency.

7. The National Lime Association. Effective Use of Lime for Flue Gas Desulfurization. Conference Proceedings, Denver, Colorado, September 27-28, 1983.

8. GAI Consultants, Inc. Coal Ash Disposal Manual, 2nd Edition CS-2049 Final Report, October, 1981. Prepared for Electric Power Research Institute.

9. Electric Power Development Co. Ltd., and Kawasaki Heavy Industries, Ltd., Study and Analysis of Fluidized-Bed Combustion Ash Disposal System, Final Report, October, 1983. Prepared for Electric Power Research Institute.

10. Battelle Columbus Laboratories. Preliminary Analysis of 6- x 6-ft AFBC Facility Emissions, EPRI CS-3188, Project 1179-8. Final Report, July, 1983. Prepared for Electric Power Research Institute.

11. U.S. Department of Energy. The Proceedings of the Sixth International Conference on Fluidized Bed Combustion, Volume III - Technical Sessions. Atlanta, Georgia, April 9-11, 1980.

12. Stone, Ralph and Richard L. Kahle. Environmental Assessment of Solid Residues from Fluidized-Bed Fuel Processing: Final Report, EPA-600/7-78-107. June, 1978. Prepared for U.S. Environmental Protection Agency.

13. Grimshaw, T. W., R. A. Minear, A. G. Eklund, W. M. Little, H. J. Williamson, and J. E. Dunn. "Assessment of Fluidized-Bed Combustion Solid Wastes for Land Disposal," Project Summary, EPA Contract No. 68-02-3103, June, 1984. Prepared for U.S. Environmental Protection Agency.

14. Envirosphere Company. Environmental Control and Assessment Plan for By-Product Utilization of Atmospheric Fluidized Bed Combustion. June, 1980. Prepared for Tennessee Valley Authority.

THERMAL OXIDATION OF SPRAY DRYER FGD WASTE PRODUCT

S. Bengtsson, Ph.D., Manager, R & D, Flakt Industri AB
S. Ahman, Staff Chemical Engineer, Flakt AB
T. Lillestolen, Manager, FGD Process Systems, Flakt, Inc.
G. Koudijs, Mechanical Engineer, Dorr-Oliver B.V.

ABSTRACT

Spray dryer FGD for coal fired boilers using low and medium sulfur coals has become an important technology. Typically, the FGD end product has been disposed of, untreated, in landfill areas.

In some countries in Europe, disposal by these means is not permitted. These countries require that the FGD end product be reclaimed for alternate use. To meet this requirement, Flakt entered into a cooperative agreement with Dorr-Oliver to develop a thermal oxidation system based upon the Dorr-Oliver FluoSolids Technology. This system provides for the calcination of the Dry FGD end product to produce a technical grade anhydrous calcium sulfate, which can be used as a setting retarder for Portland Cement.

This paper describes the process and presents operating experience and test results from a pilot plant program, and a demonstration plant (2.5 metric ton/hour) test program. Further, a description is included on how this process has been integrated into a commercial Spray Dryer FGD System recently sold by Flakt in the Federal Republic of Germany.

INTRODUCTION

Since the mid-1970's, Flakt has taken an active role in developing Dry FGD technology for the industrial and utility markets, worldwide. As a result, Flakt has successfully developed both dual-fluid nozzle atomization and mechanical rotary atomization. Additionally, Flakt has integrated the designs for dry scrubbing and particulate removal (electrostatic precipitators and fabric filters) to assure optimum and efficient Dry FGD systems.

Most recently, in order to accomodate new and changing requirements, Flakt has taken a further step in its development effort by introducing a process enhancement for treating Dry FGD waste product. The following serves to describe this development.

THE GERMAN FGD MARKET

The technologies for Flue Gas Desulfurization in the Federal Republic of Germany (FRG) have been developed to produce marketable end products such as high grade gypsum and fertilizers. The general policy among the German authorities has been to require reuse of the FGD waste product to avoid disposal. In this respect the German and Japanese market situations are very similar. To increase the market for FGD-gypsum, the German authorities have also indicated the intention to reduce future mining of natural gypsum.

Consequently, Wet FGD technologies in Germany are normally equipped with forced oxidation and gypsum washing to produce acceptable gypsum product for wall-board and plaster production.

The Dry FGD technologies in the U.S. and Europe produce a waste product which is a mixture of different calcium components which must typically be disposed of. This procedure, however, in most cases is not acceptable in Germany. Therefore, Flakt decided to investigate an alternative to disposal. This was to thermally oxidize the Dry FGD product to produce a technical grade anhydrite to be used as a setting retarder for Portland cement.

To determine whether the production of anhydrite from a Dry FGD plant was commercially viable, a cost study was made. The study compared the Flakt Dry FGD process, with thermal oxidation, and the typical wet limestone FGD process with the production of marketable gypsum, for various power plant sizes and capacity factors. The results of the study showed (Figure 1) that Dry FGD System with thermal oxidation was commercially attractive for power plants up to about 300 MW (e) for capacity factors in the range of 0.35 to 0.70. Those size and operating characteristics are typical for many German coal fired boiler plants.

The projected total production of synthetic calcium sulfate (gypsum) in the FRG in the year of 1995 is about six 106 metric tons/year. This includes five 106 metric tons/year produced by FGD installations at coal fired power plants.

The projected consumption of calcium sulfate, on the other hand, is anticipated to remain rather constant at three to four 106 metric tons gypsum/year, and about one and one-half 106 metric tons anhydrite/year.

Consequently, a total reuse of all FGD gypsum, as shown in Figure 2, will hardly be possible. However, a limited number of FGD plants producing anhydrite of acceptable quality should have good possibilities to find marketing outlets.

Based on these conclusions, Flakt and Dorr-Oliver proceeded to jointly develop a system for thermal oxidation of Dry FGD waste products to produce a reuseable technical grade anhydrite.

TECHNICAL GRADE ANHYDRITE USAGE

The technical grade anhydrite which can be produced from the Dry FGD process is a product with quite different possibilities for reuse compared to the original "sulfite" material. When mixed with water it will set due to formation of gypsum. This reaction is catalyzed by the excess lime which will slake to hydroxide.

The consumption of calcium sulfate as setting retarder in the FRG is about two 106 metric tons/year. The natural anhydrite part is 25 to 50 percent of this amount. The exact ratio between the dihydrite and anhydrite amounts are not well known since often natural mixtures are utilized.

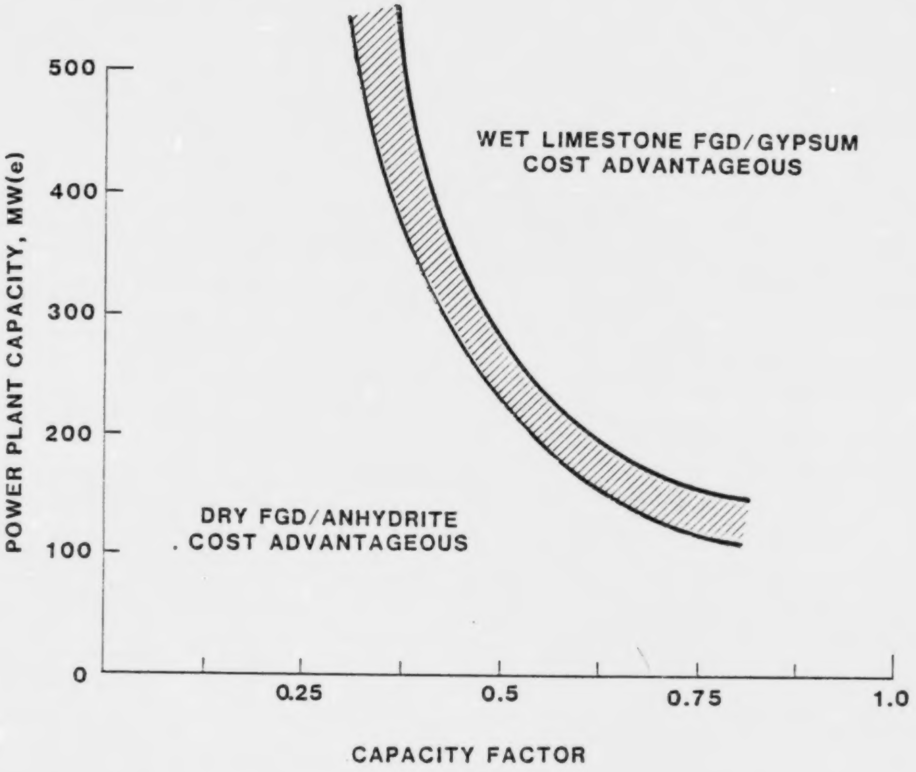


FIGURE 1. Comparison of Total Costs Between Wet and DRY FGD (No Disposal)

BASE: 1.5% Sulfur Coal, 20 Years, 8% Interest
German Retrofit Conditions

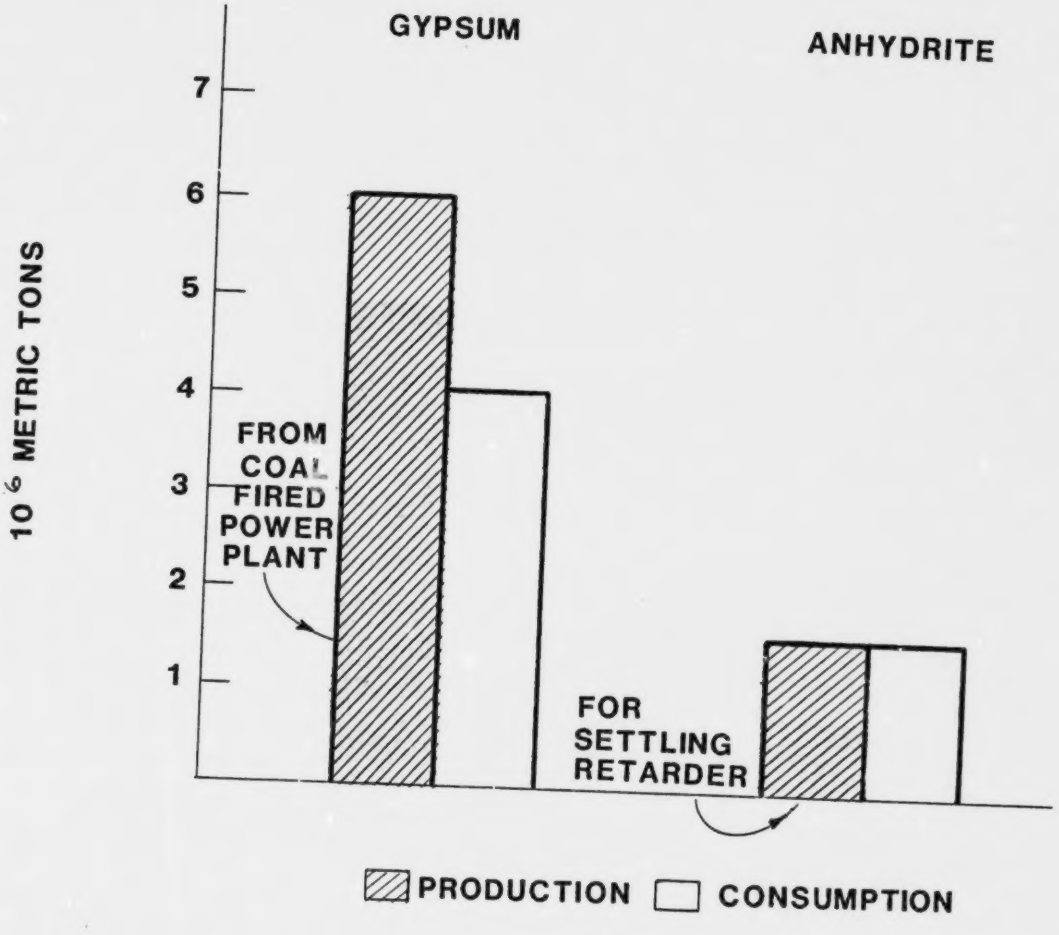


FIGURE 2: 1995 Production/Consumption Comparison for Gypsum and Anhydrite (FRG)

The use of anhydrite within the Portland cement industry has increased considerably in the last ten years, when research showed that the replacement of gypsum by anhydrite helped to solve a number of problems [1]. Typically 1/1-mixtures of gypsum and anhydrite are used as setting retarder, but some Portland cement plants use mixtures with as much as 75 percent anhydrite.

A large number of tests with technical grade anhydrite, as setting retarder for Portland cement, has been carried out by different German cement companies. A research program has also been conducted at the Technical University of Clausthal to investigate the characteristics of technical grade anhydrite utilizing a number of different German cement clinkers. All this work proves that technical grade anhydrite, as produced in the Flakt/Dorr-Oliver thermal oxidation system, is feasible as Portland cement setting retarder [2]. In fact, a considerable increase in the early strength development of cement test specimens has been determined. However, due to its chloride content, the use of technical grade anhydrite from the dry FGD process in Portland cement may have to be limited, since the DIN 1164 (German Cement Industry Standard) prescribes a maximum content of 0.1 percent Cl in the cement.

Figure 3 shows the concentration of chloride in the technical grade anhydrite that can be accepted for different setting retarder mixtures of gypsum and anhydrite assuming that the chloride contamination of the gypsum and clinker can be neglected. The results are given for 2 percent and 3 percent SO_3 content.

PROCESS CHEMISTRY

The solid reaction products from a Dry FGD plant is a mixture of different calcium components combined with some fly ash and other "inert" material originating from the lime absorbent. The material is typically very fine with particulate agglomerates ranging from 5 to 50 microns.

The chemical composition is a function of the flue gas SO_2 content and lime feed rates. A certain fluctuation of the chemical composition in a spray drying process is inherent and due to the nature of process control and equipment operation. Figure 4 shows the typical composition when firing coal of about 1% sulfur and operating the dry scrubbing plant at about 90% SO_2 removal efficiency.

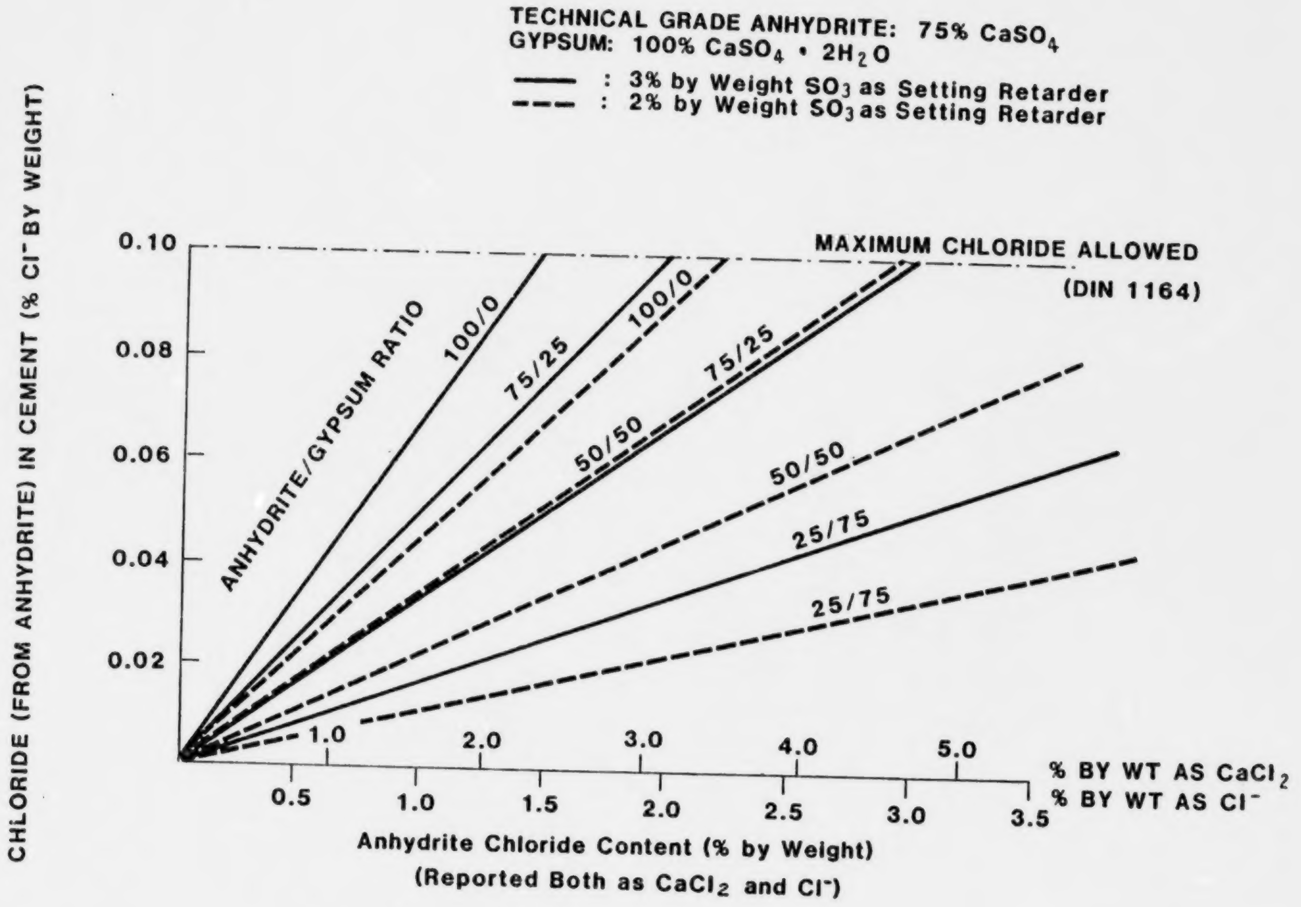


FIGURE 3: Comparison of Chloride Content in Cement to Chloride Content in Anhydrite Additive

COMPONENT	% BY WEIGHT
$CaSO_3 \cdot \frac{1}{2}H_2O$	50 TO 70
$CaSO_4 \cdot 2H_2O$	5 TO 10
$Ca(OH)_2$	5 TO 15
$CaCO_3$	2 TO 5
$CaCl_2 \cdot XH_2O$	2 TO 4
Fly Ash, SiO_2, Al_2O_3	1 TO 5
Moisture	1 TO 3

FIGURE 4: Typical Dry FGD Waste Product Composition

This "sulfite" material has so far not been found useful and is instead disposed of in landfill areas.

THERMAL OXIDATION

The chemical reactions taking place during the calcination or thermal oxidation of the sulfite material at a temperature of 800 to 850°C are summarized in Figure 5.

All crystal water will dehydrate and evaporate during the calcination.

The calcines consist of about 65 to 75% calcium sulfate (Figure 6). X-ray analysis have shown that the crystal form is equivalent to the natural B-anhydrite.

The main reaction (1) is strongly exothermic and the generated heat provides a considerable amount of the energy required for the continuous operation of the thermal oxidation. Theoretically it is quite possible to make the process autogeneous. However, in practice this would require a lower oxidation temperature range where the disproportionation of $CaSO_3$ into CaS according to

$$4CaSO_3\ (s) \longrightarrow CaS\ (s) + 3CaSO_4\ (s)$$

is considerable. Investigations reveal that the disproportionation product, CaS, has a maximum concentration when the thermal oxidation is carried out in a temperature range of 600 to 700°C (see Figure 7). This is also the temperature range where the oxidation will start to become fairly complete.

The Dry FGD product contains certain small amounts of compounds of magnesium, sodium and potassium, which will form eutectoids with melting points ranging from below 600°C to the operating temperature of 800 to 850°C. The melted phase will create a stickiness of the fine sulfite feed and cause agglomeration and formation of coarse material, pellets, in the particle size range of 0.5 to 3 mm.

This is a well-known phenomenon occurring in a fluidized bed calciner and utilized in the Dorr-Oliver proprietary processes. For example, the pelletization in a fluid bed lime mud reburning process is based on the eutectoid between Na_2SO_4 and Na_2CO_3 [3, 4].

	ΔH (KCAL/MOL)
1) $CaSO_3 \cdot \frac{1}{2}H_2O\ (s) + \frac{1}{2}O_2\ (g) \longrightarrow CaSO_4\ (s) + \frac{1}{2}H_2O\ (g)$	-57.56
2) $CaSO_4 \cdot 2H_2O\ (s) \longrightarrow CaSO_4\ (s) + 2H_2O\ (g)$	28.25
3) $Ca(OH)_2\ (s) \longrightarrow CaO\ (s) + H_2O\ (g)$	26.09
4) $CaCO_3\ (s) \longrightarrow CaO\ (s) + CO_2\ (g)$	46.62
5) $CaCl_2 \cdot XH_2O\ (s) \longrightarrow CaCl_2\ (s) + XH_2O\ (g)$	58.92 (X=4)

FIGURE 5: Thermal Oxidation Chemical Reactions

	% BY WEIGHT
$CaSO_4$	65 to 75
CaO	5 to 15
$CaCO_3$	1 to 5
$CaSO_3$	1 to 2
CaS	0.1 to 0.3
Cl	1 to 3
Fly Ash, SiO_2, Al_2O_3	3 to 10
Moisture	0 to 1

FIGURE 6: Typical Technical Grade Anhydrite Composition

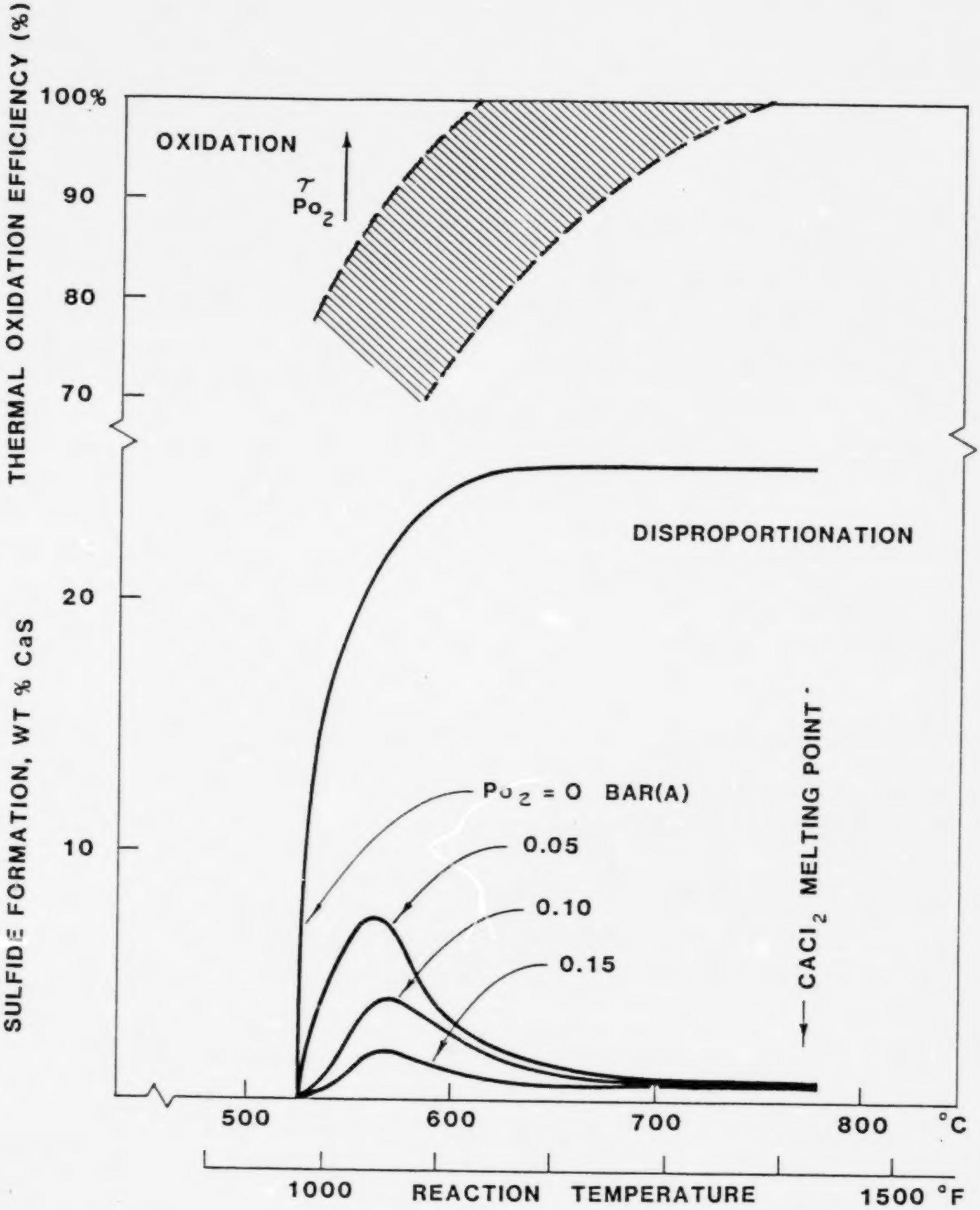


FIGURE 7: Oxidation and Disproportionation of Calcium Sulfite from Dry FGD Product at Atmospheric Pressure

(τ = Reaction Time, Po_2 = Oxygen Partial Pressure)

PILOT PLANT WORK

The thermal oxidation development work has included testing on pilot scale in different types of equipment. The most feasible technology has been found to be the fluidized bed, which has had a dominating position within the calcining and roasting operations for decades.

After early laboratory testing of the basic thermal oxidation concept in 1980-1981, the important development work was carried on by Flakt and Dorr-Oliver in the Dorr-Oliver Development Center, Stamford, CT, in 1983. Here a range of process conditions were surveyed to determine the commercial viability of fluidized bed thermal sulfite oxidation. The capacity of the Dorr-Oliver laboratory pilot plant is about 60 Kg/hour (~130 lb/hour) of solid material.

The testing showed that over 95% sulfite oxidation could be achieved at 6 to 8 percent oxygen in the temperature range of 800 to 850°C. The test work further revealed that a high concentration of fly ash in the sulfite feed impaired the pelletization and fluidbed capture. Consequently, the precollection of fly ash (described below) before the SO_2 absorption in the dry scrubbing process is essential.

The successful pilot work at the Dorr-Oliver Development Center resulted in the erection of a small commercial size demonstration thermal oxidation unit at Flakt Industri AB in Vaxjo, Sweden, Figure 8, with a capacity of 2.5 metric ton solids/hour. This unit was started-up in late 1984 and has been run batchwise since then to produce large quantities of technical grade anhydrite for full scale testing in German cement works and to demonstrate the technology for customers. The feed material is trucked from the two Flakt Dry FGD installations at Sodertalje and Malmo, in Sweden.

PROCESS DESCRIPTION

The Dry FGD process with thermal oxidation includes two important steps, which are separation of fly ash and oxidation of the product.

Reproduced from
best available copy.

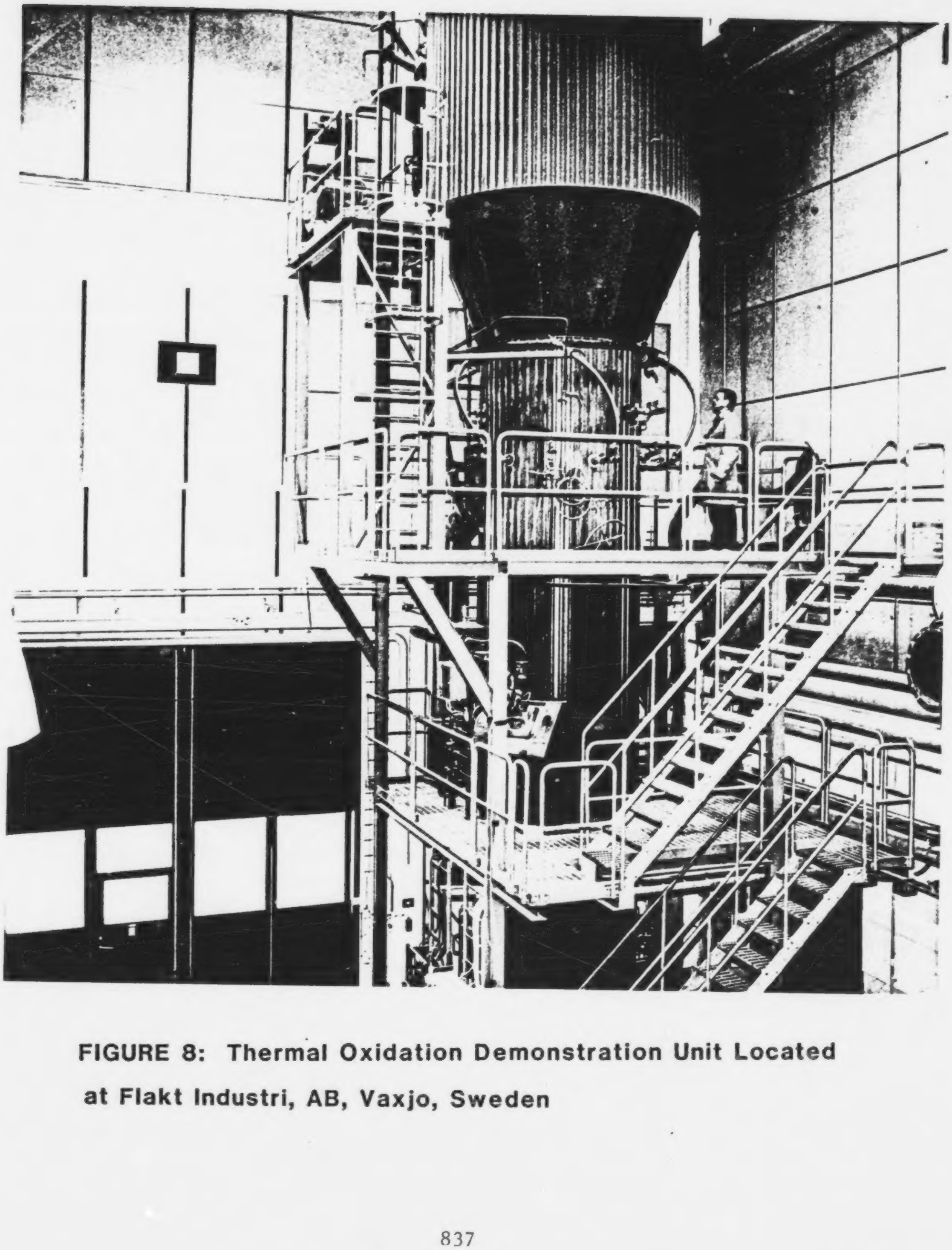


FIGURE 8: Thermal Oxidation Demonstration Unit Located at Flakt Industri, AB, Vaxjo, Sweden

SEPARATION OF FLY ASH

For the production of technical grade anhydrite it is necessary to attain a low fly ash content in the final product. Thus, a fly ash precollector is installed ahead of the spray tower.

The precollector also yields other advantages:

a) Uncontaminated fly ash. This point is of considerable importance in most European countries where fly ash is used for the production of cement and concrete. As much as about two-thirds of the fly ash production in the FRG is in fact utilized for these or other purposes.

b) Increases recirculation of more "active" reagent. Instead of recirculation of non-reactive fly ash, more lime can be taken back to the process, which means that the precollector in fact reduces the lime consumption for the Dry FGD process.

c) Less wear. Considerably less fly ash is allowed into the spray tower. The coarse fly ash would have contributed to additional wear of various process equipment (e.g., pumps, valves, etc.).

This precollector concept is included at 12 Dry FGD systems supplied by Flakt.

OXIDATION SYSTEM

The oxidation system is schematically shown in Figure 9. The heart of the oxidation system is the FluoSolids calciner (Figure 10). It comprises two fluidized beds, one (upper) in which the oxidation reaction takes place and another one (lower) which is used for cooling.

The fluidized bed consists of mainly product material, through which the upward moving gas flow is passing. By this rising gas flow the mass of solid particles are lifted and become suspended. In this way the surface of each particle is fully surrounded and in contact with the

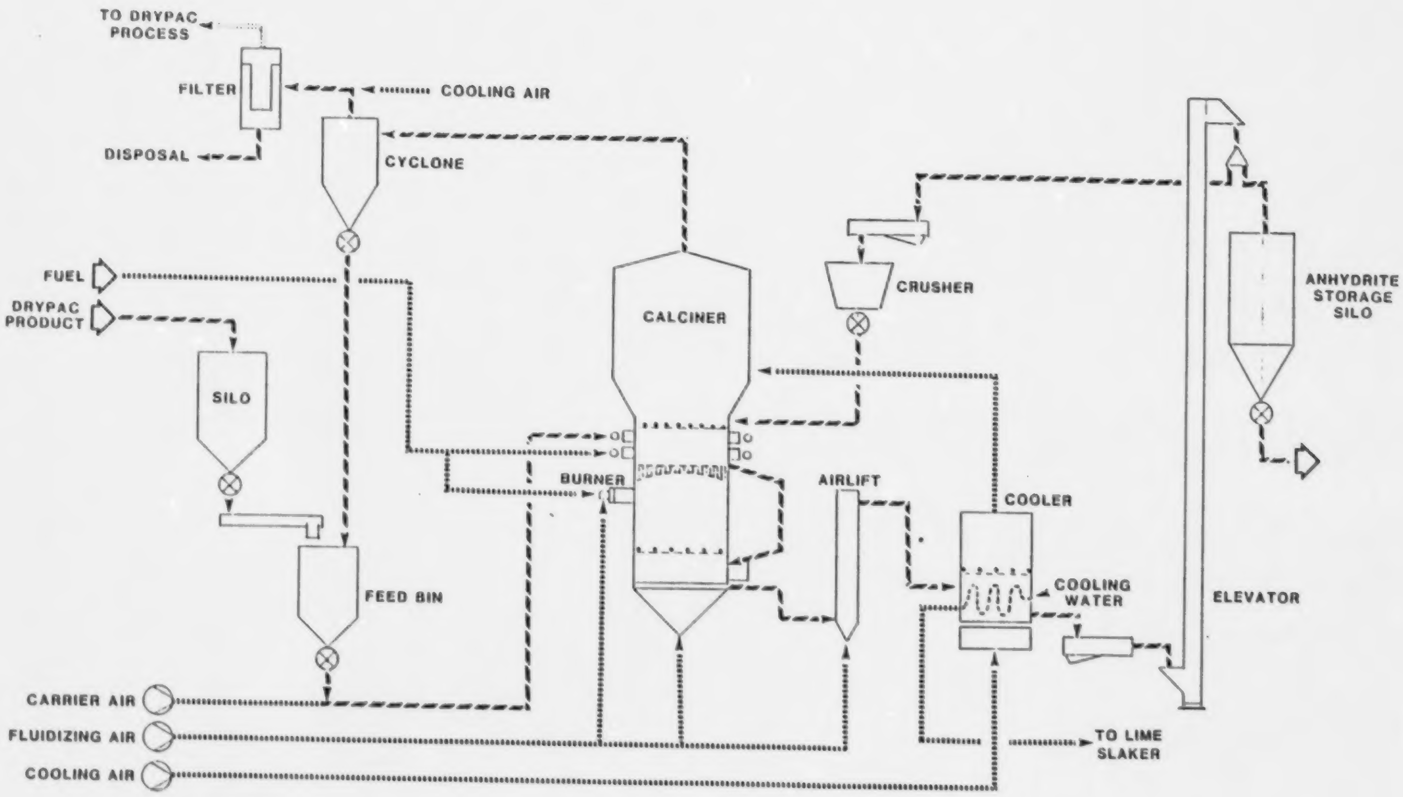

TO DRYPAC PROCESS
FILTER
COOLING AIR
DISPOSAL
CYCLONE
FUEL
DRYPAC PRODUCT
SILO
FEED BIN
CALCINER
BURNER
CRUSHER
AIRLIFT
COOLER
COOLING WATER
ANHYDRITE STORAGE SILO
ELEVATOR
CARRIER AIR
FLUIDIZING AIR
COOLING AIR
TO LIME SLAKER

FIGURE 9: Thermal Oxidation Process Flow Diagram

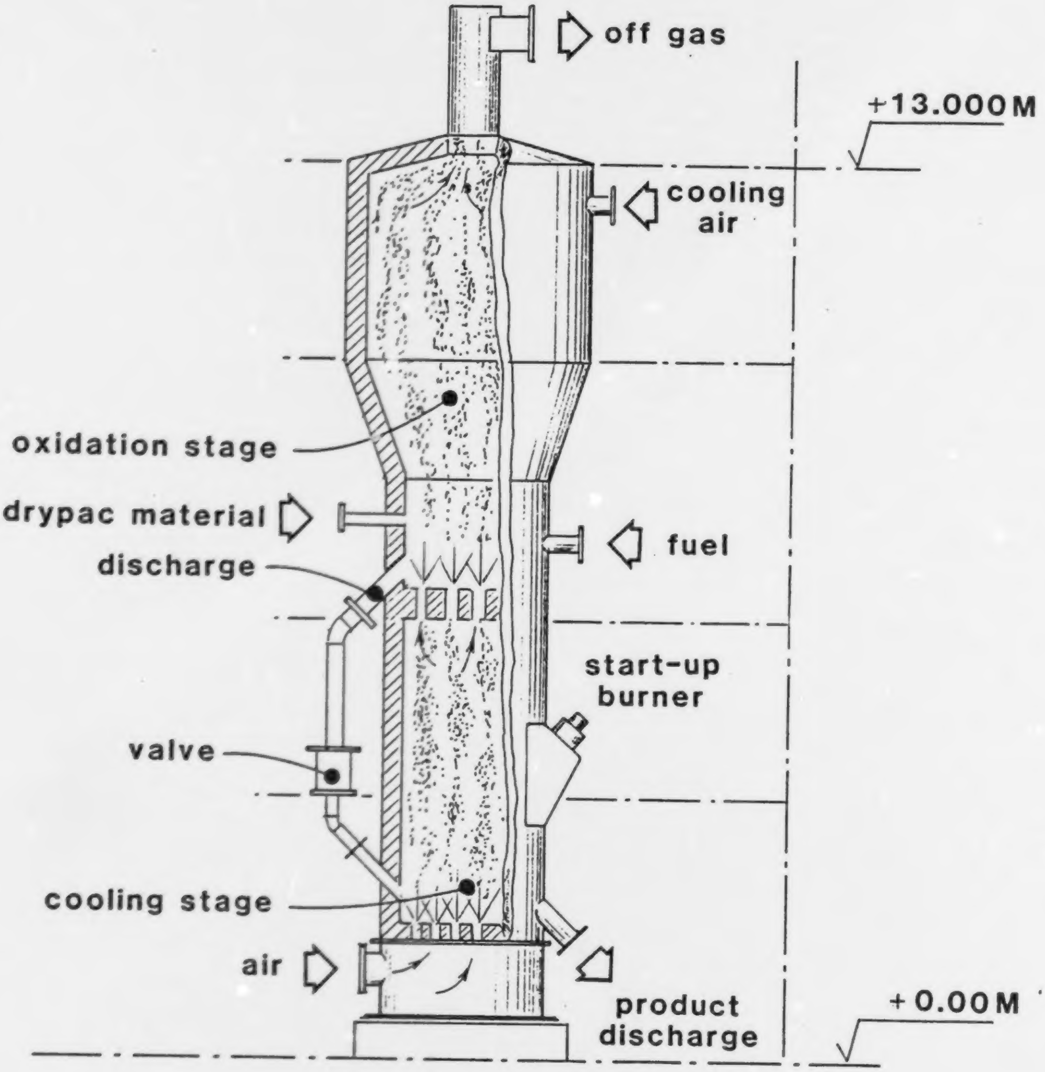

off gas
+13.000M
cooling air
oxidation stage
drypac material
fuel
discharge
start-up burner
valve
cooling stage
air
product discharge
+0.00M

FIGURE 10: Fluosolids Reactor

fluidizing medium. This solids/gas mixture behaves very much like a boiling liquid. It overflows weirs and follows the law of communicating vessels. The fluidized bed is then handled, measured, and controlled as though it were a liquid.

During fluidization, the particles are in an extreme turbulent condition, causing a random movement of the particles through the bed and producing a continuous and violent intermingling of the solids and gases. This results in an even distribution of particles and particle size as well as uniform bed temperatures, throughout the bed. This intimate contact between solids and gases also results in virtually instantaneous heat transfer between the particles and the fluidizing medium, and higher chemical reaction rates.

The waste product from the Dry FGD plant is fed pneumatically into the upper bed. The energy required for heating the product to the oxidation temperature is supplied to the bed as a fuel, e.g. by injection of oil or coal into the bed.

The oxidized product leaving the upper bed is supplied to the lower one, where the product is cooled to approximately 450°C. The oxidation air is thus preheated as the product is being cooled.

The hot air is then supplied to the upper bed. There it serves the dual purpose of fluidizing medium and as oxygen source for the oxidation and combustion processes.

The hot pellets leaving the cooling bed at a temperature of approximately 450°C are supplied to a product cooler, containing water cooled tube bundles. The pellets are then transferred to a storage bin at a temperature below 100°C.

The hot flue gases leaving the oxidizer may be cooled by dilution with ambient air or by injection of water. Finally, the flue gases pass through a particulate collector - in most cases a bag filter.

FIRST COMMERCIAL INSTALLATION

The first commercial installation of a thermal oxidation system for the production of technical grade anhydrite will be at H.K.W. - Sandreuth, Nurnberg, FRG, owned by EWAG (Energi - und Wasserversorgung AG). This is a district heating plant (see Figure 11) with three coal fired boilers each with a capacity of 106 MW_{th} [approximately 40 MW (e)].

As a result of new legislation, the "Crosskesselfeuerungsverordnung" of July 1, 1983 and the "Unweltministerkonferenz" of April 5, 1984, Sandreuth will be retrofitted with an FGD, as well as a DeNOx system.

The Dry FGD system is comprised of three spray towers/ fabric filter trains, a lime storage and preparation system, as well as a thermal oxidation stage with intermediate and final storage silos.

Milk of lime is used as reagent in the Dry FGD system which is designed to yield an SO_2 emission of less than 200 mg/Nm3 [~65 ppm (v), wet basis]. The dust emission is guaranteed to be less than 50 mg/Nm3 (~0.02 gr/SCF).

The DeNOx process to be installed will be a selective catalytic reduction (SCR) system, with ammonia as the reduction agent.

The thermal oxidation process has a capacity of 3 metric tons feed material per hour. It comprises all equipment as described previously, the FluoSolids reactor and product cooler as well as storage and feed system for the feed material and a storage silo for the technical anhydrite.

The flue gases from the FluoSolids reactor will be cleaned in a high ratio fabric filter.

The technical grade anhydrite produced will be shipped to nearby cement manufacturers.

Plant:	H.K.W. - Sandreuth
Type:	District Heating Plant
Location:	Nurnberg, FRG
Owner:	Energi und Wasserversorgung AG (EWAG)
Capacity:	3 x 106 Mw (thermal)
Sulfur Content in Coal:	0.8 - 1.2%
Gas Flow:	3 x 201,000 m^3/hr
Spray Dryer Inlet Temp:	140°C
Spray Dryer Outlet Temp:	67°C
Number of Reactors:	3
Absorbant:	Lime
Atomizer/Reactor:	4 Nozzle Heads/Reactor
SO_2 Removal:	95%
Anhydrite Production Capacity:	3 mton/hr
Contract Award:	December 1984
Start-Up:	November 1986

FIGURE 11: Sandreuth Plant Description Summary

REFERENCES

[1] Locher T. W., Richartz W., Sprung S.; Zement-Kalk-Cips 33 (1980) 271-277.

[2] Odler I., Research Report March 13, 1985 (Unpublished).

[3] Wall C. J., Graves J.T., Roberts E. J., Chem. Eng., April 14, 1975, 77-82.

[4] Bingham L. M., Tappi, 52 (1969) No. 1.

The work described in this paper was not funded by the U. S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

UNPRESENTED PAPERS

RELIABILITY PROBLEMS AND SOLUTIONS OF FGD SYSTEMS

by: D. C. Agarwal
Cabot Corporation
Kokomo, Indiana 46901

ABSTRACT

At a first glance, operation of a typical limestone FGD system seems deceptively simple. However, the first generation scrubbers of the early to mid 70's proved to be a financial and operational disaster due to metals corroding in an embarrassingly short time and non-metals failing by blistering, debonding, flaking and peeling. Various solutions were tried with only partial success. An extensive research program at various research institutions and utilities to find better materials of construction which would provide the necessary reliability in a cost effective manner, led to higher alloys up the ladder of increasing Ni-Cr-Mo alloy family and new non-metallic coatings. These too proved unreliable in many cases and what resulted was a nightmare of unscheduled shut downs, soaring maintenance costs, lost revenues and the penalties imposed from buying power from the grid, while these systems were being patched up.

In the present period of the 80's, the alloy with the highest molybdenum content of the Ni-Cr-Mo family has not only provided the necessary reliability in performance but has done so in a cost effective manner via the use of a concept known as "wallpaper concept". In this method a 1/16-inch veneer of alloy C-276 is welded onto the carbon steel or lower alloy structure. Economic analysis shows this method to provide the lowest annual cost over the design life of the plant. Case histories will be shown where this method has provided maintenance-free service for over four years and the alloy sheets still look the same as when originally installed, indicating many more years to continued reliable service.

INTRODUCTION

Acid rain is a political, economic and environmental issue in the industrially developed areas of the world and the society has to pay the cost of abating pollution - pollution in the form of sulfur oxides from fossil fired power stations. Even though the chemistry of sulfur oxide removal process is deceptively simple, monumental corrosion problems have been encountered in the various zones of an FGD (Flue Gas Desulfurization) system, such as wet/dry zone of the inlet duct, outlet ducts, stack linings, damper and damper seals, reheaters and fans. These problems have caused unscheduled shut downs, soaring maintenance costs, lost revenues and penalties imposed from buying power from the grid. Hence the need for "reliability" in performance became very critical. Field experience has shown that the use of non-metallic linings over carbon steel in critical sections of FGD systems has not been totally successful. The use of standard stainless steels and higher alloys have also either failed or suffered severe localized pitting/crevice corrosion attack. However, the alloy of the Ni-Cr-Mo family containing 16% molybdenum, commercially known as HASTELLOY® alloy C-276 has provided the necessary corrosion resistance leading to the desired performance reliability. Solid components of this alloy could not be used on a wide scale due to the massive nature of scrubber components and the costs involved. A concept known as "wallpaper concept" was developed, whereby 1.6mm (0.063-inch) thick sheets of this alloy could be applied either for a new construction or retrofit job. This concept has provided cost effective and maintenance free reliable service for over four years in FGD systems of coal-fired power plants.

HASTELLOY is a registered trademark of Cabot Corporation, 1020 West Park Avenue, Kokomo, Indiana 46901 U.S.A.

FGD SYSTEM ZONES AND THEIR CORROSION PROBLEMS

Amongst the many systems currently available, wet lime or limestone scrubbing is the most advanced and commercially used. In these systems sulfur oxides are removed by counter current spraying of flue gas with an alkaline slurry, a deceptively simple process. However, this industry in its infancy quickly realized that in this seemingly simple process, monumental corrosion problems existed. An FGD system can be separated into eight distinct regions, Figure 1, each with its own specific corrosive conditions. The major problem areas are wet/dry interface zones in the inlet to the scrubber, outlet duct beyond the outlet dampers and the stack lining. The failure of the outlet duct or stack lining will lead to immediate plant shut down and the utility is forced to purchase power to meet the demand of its customers - a very costly endeavor. A detailed description of the various zones of an FGD system is given elsewhere[1-4].

The major factors causing problems in FGD systems are sulfur content of coal, temperature, pH, chlorides, fluorides, dew point corrosion, gas velocity (erosion-corrosion, flyash erosion) closed/open loop, under deposit corrosion, upset conditions and mode of reheat. During upset, especially temperature excursions catastrophic failure of non-metallic coating/lining

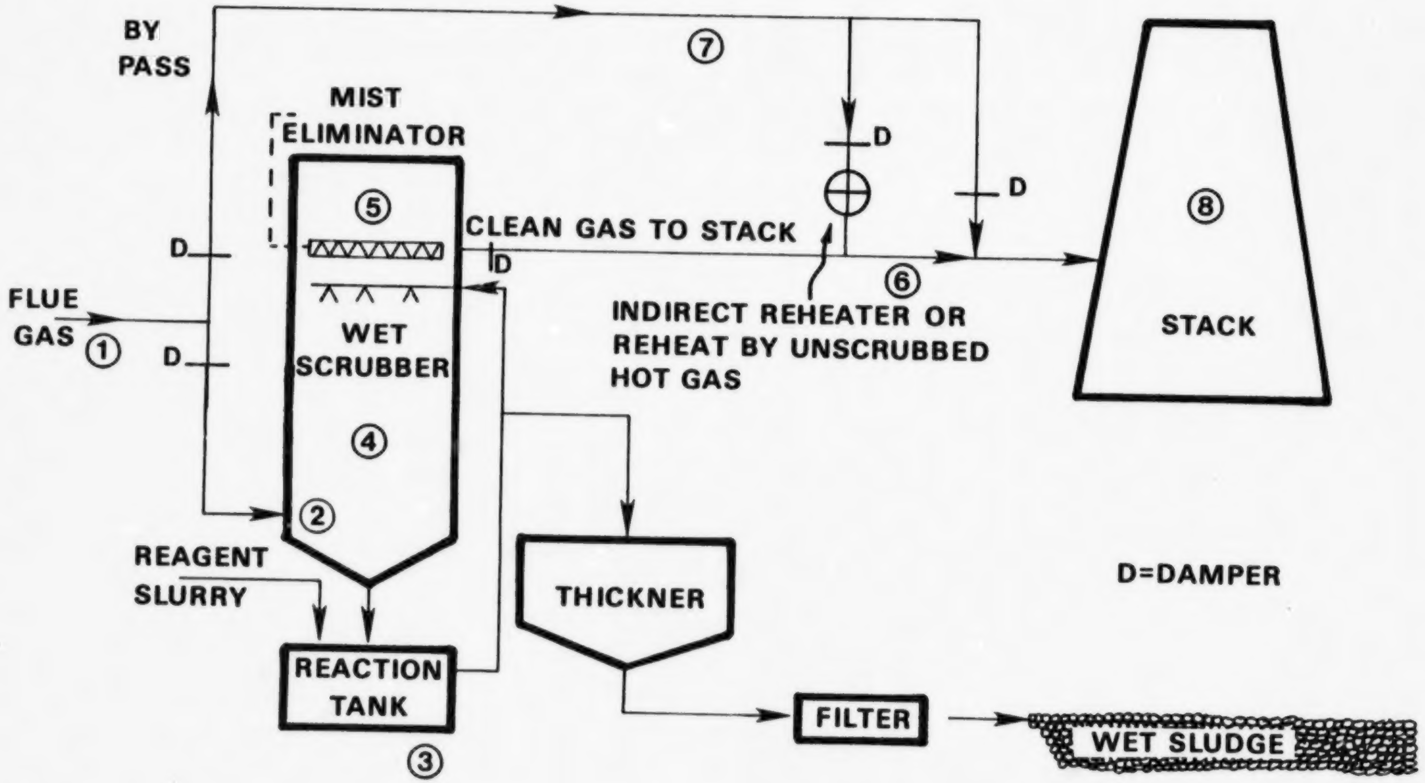

BY
PASS
⑦
MIST
ELIMINATOR
⑤
D
D
D
D
CLEAN GAS TO STACK
⑥
FLUE
GAS
①
WET
SCRUBBER
④
②
INDIRECT REHEATER OR
REHEAT BY UNSCRUBBED
HOT GAS
⑧
STACK
REAGENT
SLURRY
REACTION
TANK
③
THICKNER
D=DAMPER
FILTER
WET SLUDGE

FIG. 1: CONVENTIONAL LIMESTONE WET SCRUBBING SCHEMATIC.

can result. Most FGD systems are being operated in closed-loop mode to conserve water and reduce pollution. As a result, chloride concentrations in scrubber liquors can build up to 100,000 ppm, causing pitting and perforation of many alloys. The pH control is extremely critical and has to be maintained within narrow limits. If the pH gets too high, heavy scale deposits can form. If pH falls too low, severe corrosion results. In certain installations, part of unscrubbed hot gases are used to reheat the scrubbed cold gases. This creates very severe corrosive conditions in the mixing area and the outlet duct including the breeching. The condensates formed here in certain cases have pH values of less than one. The complexity of corrosion/erosion problems and the effects of trace elements have been addressed in many laboratory studies and field trials[5-7].

MATERIALS OF CONSTRUCTION

Since FGD systems are actually chemical plants, with different chemical and corrosive environments, materials of construction must be carefully selected. Material options presently available to the system designer/ supplier are organic linings and plastics, ceramic and other inorganic materials and metals and alloys.

Major problems occurring with non-metals are their unforgiving nature when improperly applied, plus an extreme vulnerability to upset conditions such as temperature excursions. Four major classes of metals and alloys have been considered which are used to combat the major aqueous corrosion problems. These construction materials have been discussed in greater detail elsewhere[1,8,9]. The major corrosion problem is localized attack (pitting, crevice and under deposit corrosion). Extensive laboratory and field testing[5,6,9,10] have clearly shown that the best alloy to resist these and other forms of corrosion is alloy C-276, an alloy of the Ni-Cr-Mo family, with the highest molybdenum content (Table 1). Use of this alloy in thin sheets via the wallpaper concept, is very cost effective as shown later in the "economics" section.

WALLPAPER CONCEPT - CONCERNS

Various concerns regarding this concept have been raised by. Specifically the questions relate to fabrication methods, service history, problems encountered, thermal stresses, plug weld spacing and weld dilution and cost effectiveness. The following sections will attempt to answer these concerns.

The definition of "wallpaper concept" is simply, "An application of a 1.6mm (0.063-inch) veneer of alloy sheet over a lower alloy or carbon steel substrate using alloy welding products". The process is not new. "Welded-in" liners have been successfully used in the chemical process and other industries for over 40 years. Figure 2 shows a simple schematic of this method.

CASE HISTORY

The aforementioned concept has already been successfully used at two major utilities in U.S.A. and other utilities are seriously considering this

TABLE 1. NOMINAL CHEMICAL COMPOSITIONS

Alloy	Ni	Fe	Cr	Mo	W	Cu	C	Others
Austenitic Stainless Steels								
316	12.0	Bal	17.0	2.0	-	-	0.08*	
317L	13.0	Bal	19.0	3.5	-	-	0.03*	
317LM	14.0	Bal	19.0	4.25	-	-	0.03*	
20-Type "Super" Stainless Steels								
20CB-3® alloy	34.0	Bal	20.0	2.5	-	3.5	0.06*	Cb-1.0*
Alloy No. 904L	25.0	Bal	19.5	4.5	-	1.5	0.02	
Alloy 700	25.0	Bal	21.0	4.5	-	0.5*	0.015	
Duplex (austenitic/ferritic) "Super" Stainless Steels								
FERRALIUM® alloy 255	5.0	Bal	25.5	3.5	-	1.7	0.03	N-0.2
Nickel-Base Alloys								
Alloy No. 825	Bal	30.0	21.5	3.0	-	2.0	0.03	
Alloy G	Bal	20.0	22.0	6.5	0.5	2.0	0.02	Cb+Ta-2
Alloy G-3	Bal	20.0	22.0	7.0	0.5	2.0	0.015*	Cb+Ta-0.3
Alloy No. 625	Bal	2.5	21.5	9.0	-	-	0.05	Cb+Ta-3.5
Alloy C-276	Bal	5.0	16.0	16.0	4.0	-	0.02*	V-0.35*

* Maximum

WALL PAPER CONCEPT

1/16" (1.6 mm) THICK Alloy C-276 OVER CARBON STEEL USING Alloy C-276 WELDING PRODUCTS

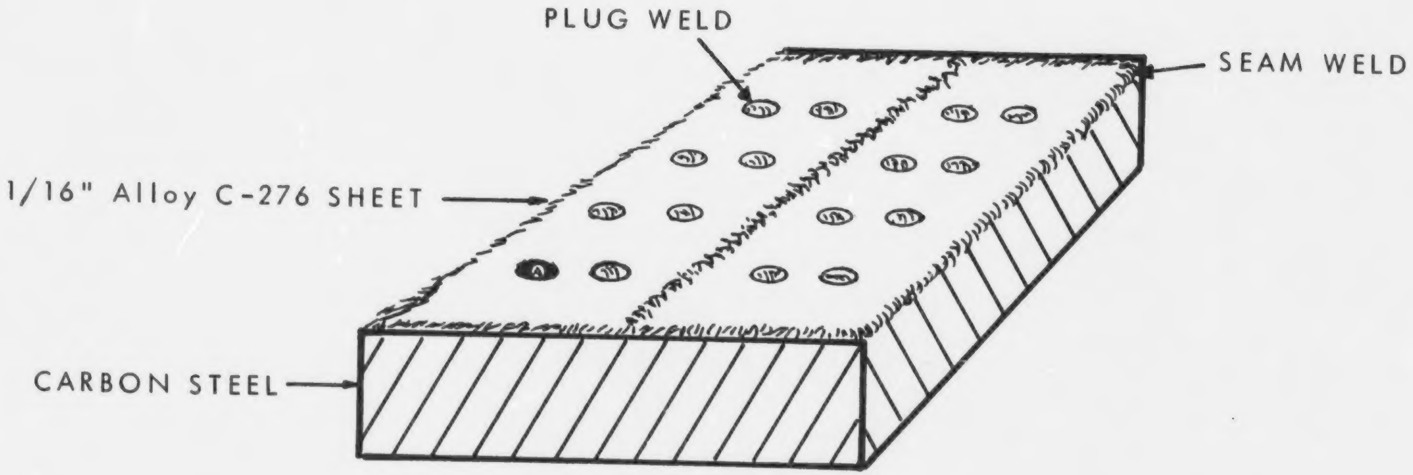

PLUG WELD
SEAM WELD
1/16" Alloy C-276 SHEET
CARBON STEEL

FIG. 2: SCHEMATIC OF WALLPAPER CONCEPT

concept to provide maintenance free service and drastic reduction of their annual maintenance budget. In fact, recently an additional four major utilities have specified this concept to solve their corrosion and reliability problems.

S. MISSISSIPPI POWER ASSOCIATION - R.D. MORROW STATION

A little over four years ago the inlet duct floor where a wet-dry interface exists and part of the mixing zone in the outlet duct where hot unscrubbed gases and moisture laden scrubbed gases are combined were lined with thin sheets of alloy C-276[11,12]. See Figure 3. A recent four year inspection showed the alloy continuing to give excellent performance. Extensive ultrasonic measurements on the alloy C-276 sheet gave consistent readings of 1.6mm ±0.025mm (0.063-inch ±0.001-inch) indicating no corrosion. There was no evidence of any localized attack either in the base metal or the weldments. Approximately 512m^2 (5500 sq. ft.) of area has been lined to date with additional areas to be lined in near future. It is felt by the utility management and corrosion engineers that alloy C-276 would continue to give maintenance-free service for many more years to come.

CENTRAL ILLINOIS LIGHT COMPANY - DUCK CREEK STATION

An alloy G outlet duct, 6.1m (20-feet) in diameter by 30.5m (100-feet) long had been used to transport scrubbed gases. When the scrubbers were not in operation it was used as a bypass duct with a water spray system to cool the flue gases from 177°C (350°F) to 121°C (250°F) prior to entering the stack. This alloy G duct had severe pitting and perforations. After two years of exposure in this duct, a test panel of alloy C-276 was free from pitting. As a result of this test and excellent performance of alloy C-276 at the R.D. Morrow plant, CILCO lined the entire duct with 1.6mm (0.063-inch) thick alloy C-276 sheet. Approximately 697m^2 (7,500 sq. ft.) of area was lined.

The plant completed one year of operation in May 1984. No evidence of any localized attack was detected, either on the base metal or weld metal. The alloy C-276 lining over carbon steel in the breeching area, was also free from any localized attack in the weld areas. A concern that the weld metal dilution from carbon steel could lower the alloy content to the extent that localized attack could occur, however, this was not the case. The welds in this area, also, were free from any attack. Ultrasonic measurements of the sheet thickness gave consistent readings of 1.6mm (0.063-inch) ±0.025mm (0.001-inch), indicating a very minimal-to-zero uniform metal loss. A single U-bend made out of 1.6mm (0.063-inch) sheet acting as the expansion joint between the outlet duct and the breeching was also free from any attack. The fabrication details at CILCO are provided elsewhere[13,14].

THERMAL STRESS

Figure 4 shows the computer calculated temperature profile and the resultant thermal stress on the weldments at the carbon steel/alloy C-276 interface. This stress is far below the yield or shear strength and hence is not a problem.

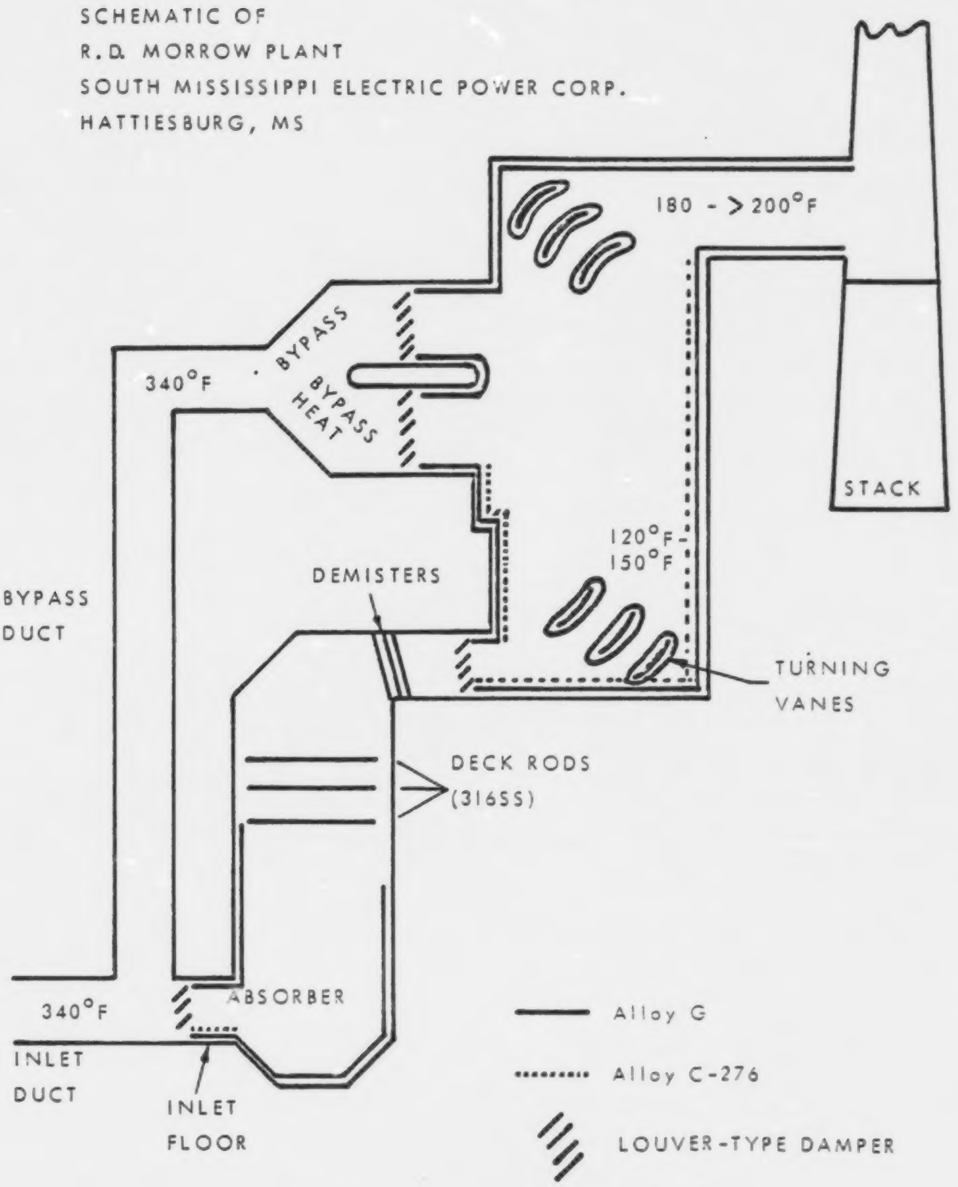
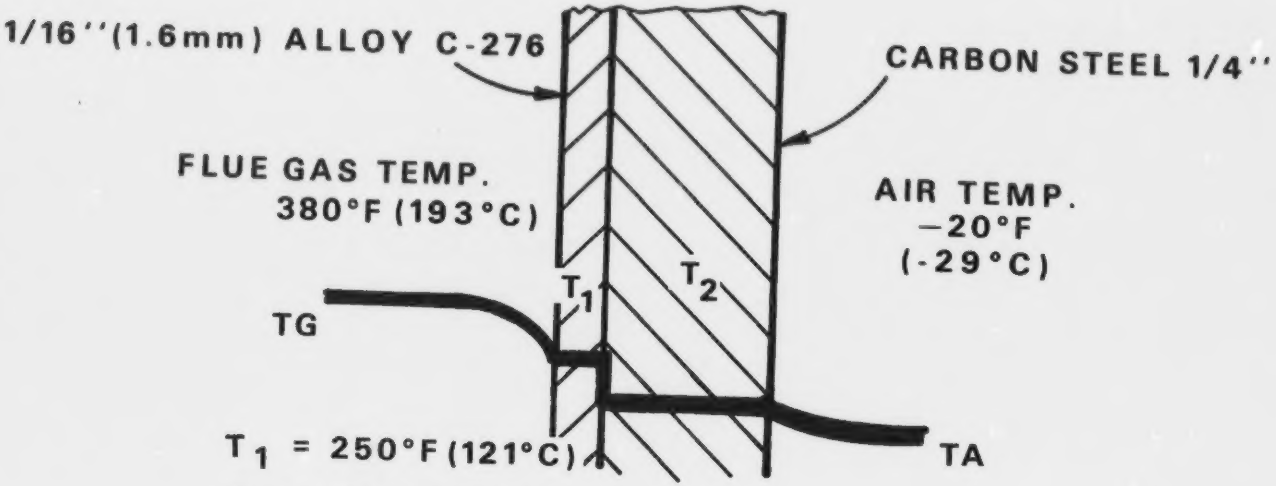

SCHEMATIC OF
R.D. MORROW PLANT
SOUTH MISSISSIPPI ELECTRIC POWER CORP.
HATTIESBURG, MS
180 - > 200°F
BYPASS
340°F
BYPASS HEAT
STACK
120°F - 150°F
DEMISTERS
BYPASS DUCT
TURNING VANES
DECK RODS (316SS)
340°F
ABSORBER
INLET DUCT
INLET FLOOR
Alloy G
Alloy C-276
LOUVER-TYPE DAMPER

FIG. 3: SCHEMATIC OF R.D. MORROW PLANT

WALL PAPER CONCEPT
CALCULATED THERMAL STRESSES


1/16''(1.6mm) ALLOY C-276
CARBON STEEL 1/4''
FLUE GAS TEMP.
380°F (193°C)
AIR TEMP.
−20°F
(-29°C)
T_1
T_2
TG
TA

T_1 = 250°F (121°C)

T_2 = 174°F (79°C)

CALCULATED THERMAL STRESS ≈ 13 KSI (89.6MPa)
TYPICAL 0.2% Y. STRENGTH ≈ ⩾ 41 KSI (286.4MPa)
TYPICAL SHEAR STRENGTH ≈ ⩾ 60 KSI (413.7MPa)

FIGURE 4: THERMAL STRESS CALCULATIONS

PLUG WELD SPACING

The pressure inside an FGD duct work is normally positive, but in certain cases a negative pressure up to 38cm (15-inches) of water can develop. In such cases the plug weld would experience stresses which try to shear the plugs. Calculations showed that even for a 122cm (48-inch) distance between plugs, the stress would not exceed the yield or shear strength of the welds. Figure 5 shows the calculated stresses for various plug weld spacings.

WELD DILUTION

Concerns have been expressed regarding the weld dilution and corrosion resistance of weldments when applying 1.6mm (0.063-inch) thick alloy C-276 over carbon steel using alloy C-276 welding products. The case history of R.D. Morrow Station and CILCO Duck Creek Station should alleviate this concern. This is further shown schematically in Figure 6. Even though the alloy G base metal pitted badly, all the welds made with alloy C-276 weld products were free from any localized attack. Figure 7 illustrates the same point, except that in this case, the plug weldment was even richer in alloy content, since it consisted of alloy C-276 sheet over carbon steel, using alloy C-276 welding product. Figure 8 shows a comparison of these two processes.

FGD ECONOMICS

FGD economics dictates the need for reliability. Installation of an FGD system may run between 20 to 25 percent of the total plant cost. This investment bears no financial return and hence the need for reliability in performance. However, with any increased reliability there is an increased associated cost. These two seemingly contradictory features of an FGD system, i.e., reliability and cost effectiveness have been reconciled by the use of wallpaper concept with alloy C-276 sheet. Table 2 gives the estimated installed cost in $/Sq. Ft. for retrofits of existing structures. On new construction, these costs using metal and non-metals are shown in Table 3.

Field data also exist showing that some non-metallic linings, which are lower in initial cost than alloys, have not performed well in the outlet duct and stack liners. These linings have either been repaired frequently or completely replaced in periods of less than six months to a few years. The recent proven alternative to the functional unreliability of even the more expensive non-metallic linings, and cost prohibitiveness of solid alloys, is the concept of lining with thin sheet of the higher alloy. This concept, if properly followed, will provide the required reliability with minimal maintenance[8].

ANNUAL COST CALCULATIONS

Many techniques exist for economic appraisal of alternative corrosion control methods. NACE Standard RP-02-72 gives a method of calculating the

WALL PAPER CONCEPT
PLUG WELD SPACING

NEGATIVE PRESSURE ≈ 5 INCHES OF WATER ≈ 0.6 PSI (4.19 kPa)

12'' SPACING → STRESS ON 1/4'' PLUG WELD ≈ 2 KSI (13.8 MPa)

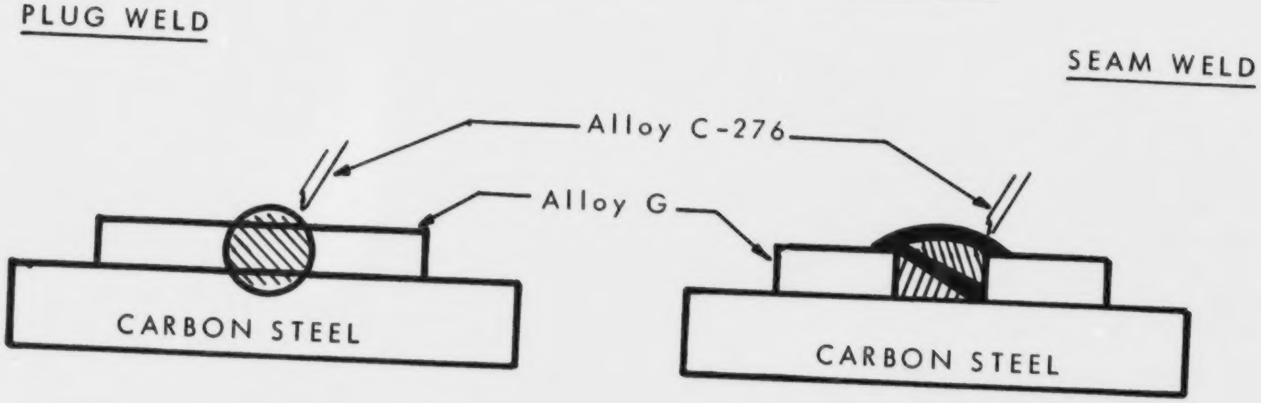
24'' SPACING → STRESS ON 1/4'' PLUG WELD ≈ 7 KSI (48.2 MPa)

48'' SPACING → STRESS ON 1/4'' PLUG WELD ≈ 28 KSI (193 MPa)

TYPICAL SHEAR STRENGTH ≥ 60 KSI

FIGURE 5. PLUG WELD SPACING CALCULATIONS

WALL PAPER CONCEPT
CONCERNS REGARDING WELD DILUTION

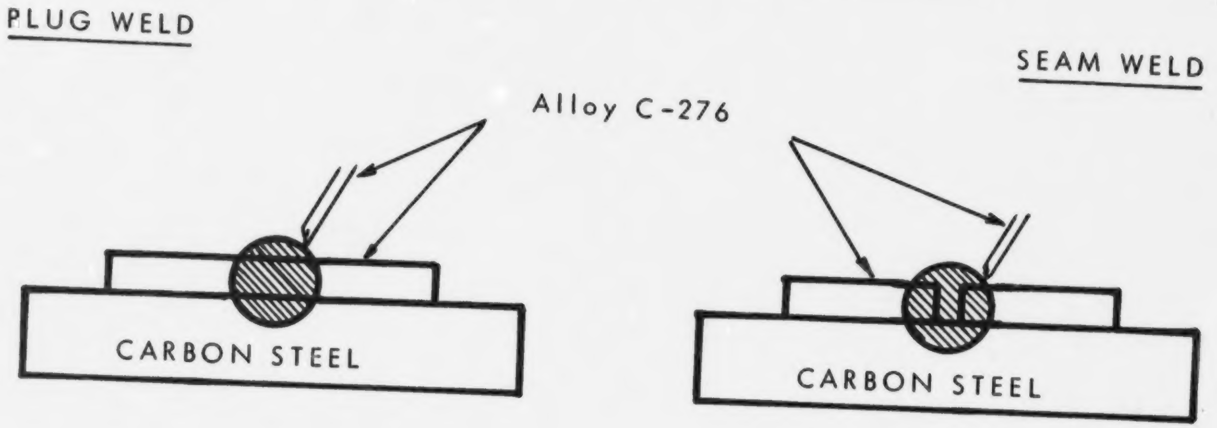

PLUG WELD
SEAM WELD
Alloy C-276
Alloy G
CARBON STEEL
CARBON STEEL

R. D. MORROW STATION

- Alloy G PITTED IN BASE METAL IN INLET FLOOR WET/DRY ZONE AND OUTLET DUCT
- NO ATTACK ON ANY WELDS
- PITTING RESISTANCE AS OBSERVED IN SERVICE
  C.S. $\lll$ Alloy G $\ll$ WELD METAL $\lesssim$ Alloy C-276

FIG. 6: WELD DILUTION CONCERNS

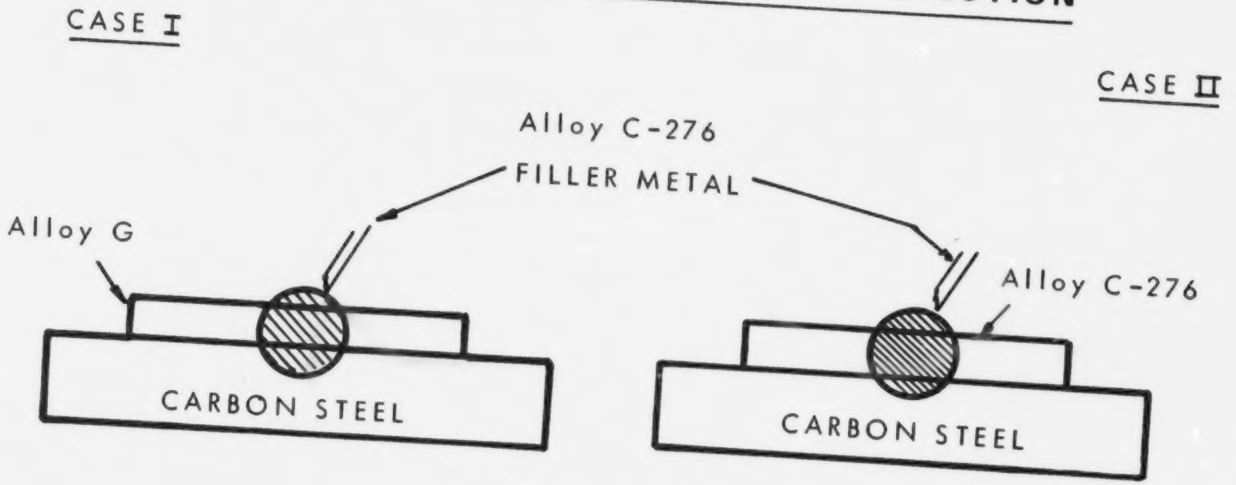
WALL PAPER CONCEPT
CONCERN REGARDING WELD DILUTION


PLUG WELD
SEAM WELD
Alloy C-276
CARBON STEEL
CARBON STEEL

AT CILCO, DUCK CREEK (BREECHING AREA) THE WELDS ARE FREE FROM ATTACK AFTER ONE YEAR OF SERVICE

FIG. 7: WELD DILUTION CONCERNS

WALL PAPER CONCEPT
CONCERN REGARDING WELD DILUTION


CASE I
CASE II
Alloy C-276
FILLER METAL
Alloy G
Alloy C-276
CARBON STEEL
CARBON STEEL

THE CORROSION RESISTANCE OF THE WELDMENT IN CASE II WILL BE BETTER THAN CASE I.

PITTING RESISTANCE

DECREASING ←——— ———→ INCREASING

$\lll$ C. S. $\lll$ Alloy G $\ll$ WELD CASE I $\leq$ WELD CASE II $\approx$ Alloy C-276

FIG. 8: WELD DILUTION CONCERNS

TABLE 2 - Retrofit Cost Comparison
Wallpaper Concept vs. Non-Metallics

Estimated Installed Cost in $/Sq. Ft.
Retrofit on Existing Structure

Alloy C-276	Alloy No. 625	Fluoroelastomer/Penn Gaurd Block Non-Metallics
$40-45	$39-44	$40-45

Cost include:
- Sheet costs
- Alloy C-276 weld wire costs
- Surface Preparation costs
- Fitting costs
- Welding costs
- Inspection and repair costs

TABLE 3 - New Construction
Estimated Installed Cost
Comparison $/Sq. Ft.

Metals Versus Non-Metals
New Construction

Method	Estimated Installed Cost
Carbon steel + CEILCOTE 871X or CXL 2000 or PENNGUARD block	$70-75
Carbon steel + neoprenre	$55-60
Carbon steel + PLASITE 4005	$50-55
Carbon steel + Con-Chem 206/207	$40-45
Carbon steel + wallpaper concept with 1/16" alloy C-276	$70-75
Clad steel with alloy 625	$75-80
Solid 1/4" alloy 625	$130-140
Solid 1/4" alloy C-276	$135-145

equivalent uniform "Annual Cost", which is well recognized and has been widely used in the industry.

Tables 4 and 5 show annual cost calculations, using the aforementioned NACE standard and certain assumptions of life, interest rate (cost of capital), tax rate, depreciation for both non-metals and the wallpaper concept. The calculations clearly show that for non-metals, the annual costs could be greater than $200K for a two-year life to approximately one-fourth the cost for the 20-year life of the wallpaper concept (Table 6). If one takes into account the lost revenues due to the unscheduled shut downs and reliability aspects, the use of alloy C-276 looks even more attractive and desirable. The excellent performance at R.D. Morrow plant gives credibility and reinforces the cost effectiveness of this method. Figure 9 graphically compares the annual cost of "wallpaper concept" alloy C-276 metal lining with non-metallic linings.

FUTURE TRENDS IN MATERIAL SELECTION

With the cost of scheduled and unscheduled maintenance going up and considering the increased need for functional reliability, materials scientists, corrosion and pollution control engineers will be paying very close attention to materials of construction. Over the last four to five years, the industry has identified materials degradation problems in these systems and has begun to develop solutions. Materials technology, as applied to scrubber environments, has graduated from infancy to adolescence and with more research will soon attain maturity.

Judicious selection of corrosion-resistant alloys will help keep downtime to a minimum. Non-metallic coatings and liners also have a significant role to play in keeping systems operational with the lowest maintenance dollars, but they do have limitations. Thus the optimum construction of future systems will have to consist of a properly selected combination of various alloys, metallic and non-metallic linings/coatings.

CONCLUSIONS

Laboratory and field experience confirm that the best corrosion resistance in critical sections of FGD is provided by the highest molybdenum containing alloy of the Ni-Cr-Mo family, alloy C-276.

This alloy has provided functional reliability in a cost effective manner, via the use of wallpaper concept.

The annual cost comparison clearly shows that wallpaper concept utilizing 1.6mm (0.063-inch) sheet of alloy C-276 provides the lowest cost over the projected life of the FGD system.

This wallpaper concept is applicable to both new construction and retrofits and can easily be applied in both shop and field environments.

Alloy C-276 is providing maintenance-free performance at many utilities.

TABLE 4 - New Construction

Wallpaper Concept Annual Cost Calculations
NACE Standard RP-02-72

"Wallpaper concept" with 0.063-inch (1.6mm) alloy C-276 sheet over carbon steel

Assumptions:

Area = 10,000 sq. ft.
Estimated installed cost = $70.00/sq. ft.
Rate of return after taxes = 10-percent
Tax rate = 48-percent
Depreciation method = Sum of digits

	10 Yr. Life	20 Yr. Life
Annual cost	$73,000	$53,000

TABLE 5 - New Construction

Non-Metals Annual Cost Calculations
NACE Standard RP-02-72

Carbon Steel with Non-Metallic Lining

Assumptions

Area = 10,000 sq. ft.
Estimated installed cost = $70.00/sq. ft.
Rate of return after taxes = 10-percent
Tax rate = 48-percent
Depreciation = Sum of digits method

Annual Cost	2 Yr. Life	5 Yr. Life
No maintenance	$224,000	$114,000
With maintenance	$242,000	$132,000

TABLE 6 - Annual Cost Comparisons

Metals Versus Non-Metals

Non-Metals*	2 Yrs.	5 Yrs.	10 Yrs.	20 Yrs.
Without maintenance	$224K	$114K	-	-
With maintenance	$242K	$132K	-	-
Metals**				
Wallpaper concept with 0.063-inches HASTELLOY alloy C-276	-	-	$73K	$53K

* Non-metals have average life between two to five years. In some instances, the life has been only a few months.

** The wallpaper concept life with alloy C-276 is expected to be greater than 10 years.

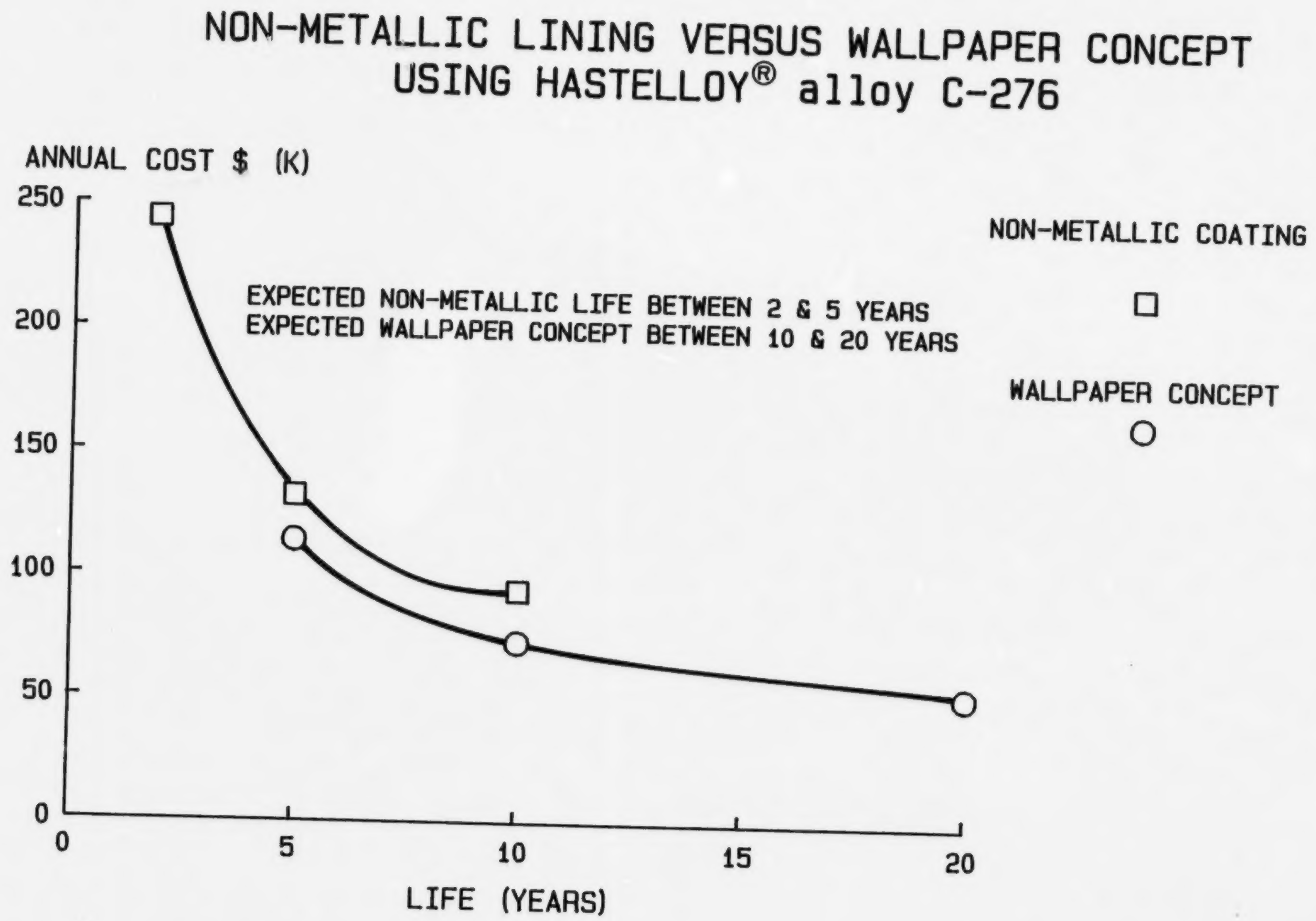


FIGURE 9: ANNUAL COST CALCULATION

Part of the work described in this paper is under copyright by NACE who have given their written permission for publication. Furthermore, the work described in this paper was not funded by the U.S. Environmental Protection Agency; therefore, the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

REFERENCES

1. Rosenberg, H. S. and others, EPRI Report NI, CS-1736, Volumes 1 and 2, March 1981.
2. Ostroff, N. and others, Proceedings, 1984 Air Pollution Seminar, Orlando, Florida, December 1984, Paper #18.
3. Crow, G. L., Horsman, H. R., Corrosion/80, Paper No. 274.
4. Maiya, P. S., Argonne National Laboratory Report, ANL-80-59, Oct. 1980.
5. Koch, G. H., Beavers, J. A., EPRI Report CS-2537, August 1982.
6. Koch, G. H. and others, EPRI Report RP-1871-2, 1982.
7. Thompon, N. G. and others, Corrosion/84, Paper No. 297.
8. Koch, G. H. and Rosenberg, H. S., Flue Gas Symposium Record, Ottawa, Canada, September 19-20, Canadian Electrical Association.
9. Silence, W. L. and others, Proceedings 1981 Solving Corrosion Problems in Air Pollution Control Equipment, Denver, Colorado, Paper #18, August 1981.
10. Dille, E. R., Froelich, D. A., Power, August 1983.
11. Agarwal, D. C., Cabot DIGEST, January 1983, Vol. 34, No. 1.
12. Agarwal, D. C., Cabot DIGEST, April 1985, Vol. 36, No. 3.
13. Agarwal, D. C., Cabot DIGEST, August 1984, Vol. 35, No. 4.
14. Agarwal, D. C., Corrosion/85, Paper No. 46.

FOUR CORNERS 4 AND 5 WASTE PROCESSING SYSTEM:
AN INTEGRATED AND UNIQUE RETROFIT

AUTHORS

Alexander P. Simko
Project Engineering Manager
Emission Abatement Projects
ARIZONA PUBLIC SERVICE COMPANY
2216 W. Peoria Avenue
Phoenix, Arizona 85036

Richard R. Lunt, Ph.D.
Project Engineering Manager
UNITED ENGINEERS AND CONSTRUCTORS, INC.
30 South 17th Street
Philadelphia, Pennsylvania 19101

Charles S. White
Sales Manager
CONVERSION SYSTEMS, INC.
115 Gibraltar Road
Horsham, Pennsylvania 19044

ABSTRACT

The Arizona Public Service Company has retrofit Particulate and Sulfur Dioxide Control Systems to Units 4 and 5 of its Four Corners Steam Electric Generating Station. As part of the SO_2 control system, APS installed a Conversion Systems, Inc. Waste Processing System to process the fly ash and FGD sludge. This System is unique in that it is:

- The latest operating retrofit in the United States;
- the largest retrofit (1600 MW) in the United States;
- a specially designed vertical layout waste processing plant for direct loading of trucks;
- designed for five modes of waste processing;
- designed to be operated on an intermittent basis; and
- designed to work with, rather than replace the existing fly ash handling system.

This paper discusses the design criteria, unique features, and initial operation of the Waste Processing System. It presents the challenges of retrofitting a large scale system to an operating power plant and the solution to these challenges selected for the Four Corners site.

BACKGROUND

The Four Corners Steam Electric Generating Station, operated by Arizona Public Service Company (APS) of Phoenix, Arizona, is located approximately 20 miles southwest of Farmington, New Mexico. In 1979, APS was required to upgrade the existing Particulate Control System, and install SO_2 removal equipment on Units 4 and 5 of the Four Corners Station, which are 800 MW each and burn New Mexico subbituminous coal. The Emission Abatement Project, as this retrofit was designated by APS, was subdivided into two projects: the Particulate Removal Project (PRP) and SO_2 Removal Project (SO_2 RP). The PRP involved installation of fabric filters to replace the existing precipitator. The SO_2 RP involved installation of a conventional lime slurry scrubbing system and the Waste Processing System. United Engineers Constructors, Inc. (UE&C) of Philadelphia, Pennsylvania was retained as the Architect Engineer and Construction Manager for the Emission Abatement Project.

Conversion Systems, Inc. (CSI) of Horsham, Pennsylvania was awarded a contract by APS/UE&C in December, 1981, to engineer, fabricate, furnish, deliver and erect a Waste Processing System for the treatment of the fly ash from the PRP project and sludge from the Wet Lime Flue Gas Desulfurization System (Figure 1). CSI had earlier performed a sludge stabilization demonstration at the Four Corners site in 1974-1975 in conjunction with testing of a prototype horizontal scrubber.

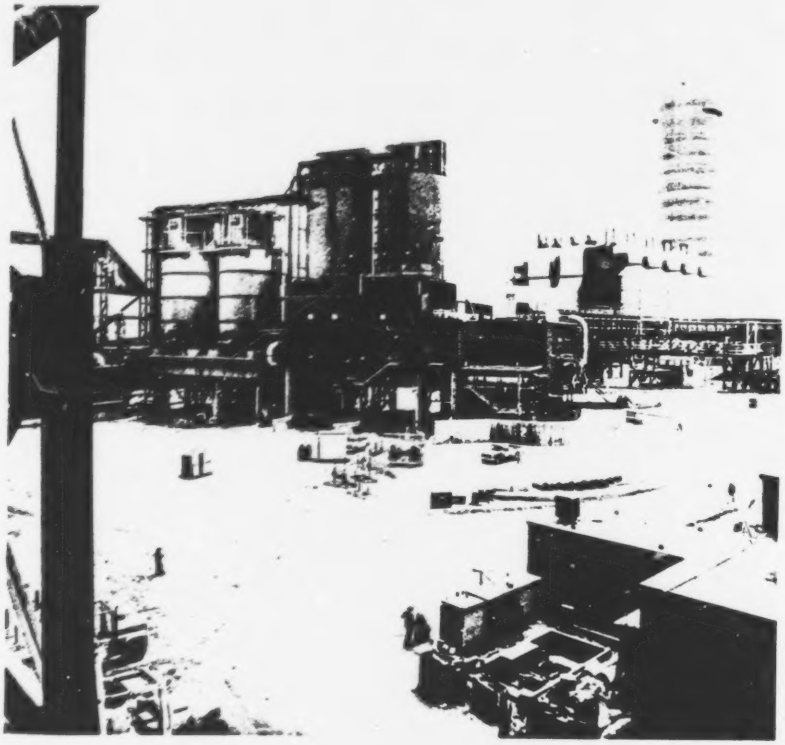


Figure 1 The Waste Processing System has been retrofit into a comparatively tight area and incorporated with the existing ash system.

SYSTEM DESIGN CRITERIA

The Waste Processing System is designed to produce an environmentally acceptable and physically stable solid waste product from the fly ash and FGD sludge produced as a result of the Emission Abatement Project. The Four Corners Generating Station is located approximately 5,300 feet above sea level. The power plant site experiences wide fluctuations in daily and annual temperature. The extreme daily temperatures range from minus 34°F to 104°F. The mean annual temperature at the site is 51°F.

Units 4 and 5 burn a maximum of 864 tons per hour of coal, combined. The coal has a typical ash content of 22% and a typical sulfur content of 0.78%. The design ash range is 19 to 25%. The design sulfur range is 0.6 to 1.0% (Table 1).

TABLE 1

COAL PROXIMATE ANALYSIS

	Content (Wt %)	
	Typical	Range
Fixed Carbon	33.6	32-35
Volatile	31.4	30-33
ASH	22.8	19-25
Water	12.2	11-13
Sulfur	0.87	0.6-1.0
Chloride	-	0.25(Max)
Fluoride	-	0.01(Max)
HHV (BTU/lb)	8600	8200-9100

The fly ash collected by the Fabric Filter System ranges between 131 tons per hour and 172 tons per hour, with an average of 152 tons per hour. Total sludge production from the lime scrubbers varies between 19 and 31.6 tons per hour with an average of 23.6 tons per hour (dry weight basis). Therefore, the ash to sludge ratio has a range of 4:1 to 9:1 (dwb).

The Waste Processing System is designed to be operated two 8 hour shifts per day. Considering start-up, shut-down, and clean-up, the actual sytem operating design time is 13 hours per day. In other words, the Waste Processing System is designed to process the daily waste generated in 24 hours within a 13 hour waste processing operating period.

The Waste Processing System consist of three parallel equal size trains (Table 2). Each train has the capacity of processing 100% of the waste fly ash and FGD sludge from one boiler, with one processing train as a common spare.

TABLE 2

PRIMARY EQUIPMENT

Item	Quantity	Description
Processing Train	3	- 2 Operating, 1 spare - 13 Hr. Design Operation - Unit cross connect - 5 Operating modes - Intermittent Operation
Vacuum Filters	3	- 12 ft. diameter X 20 ft.long - Rubber lined carbon steel wetted parts
Pugmill Mixers	3	- 7.5 ft. wide X 16 ft. long X 3.25 ft. high with 44 in. diameter paddles - Epoxy coated carbon steel shell and cover - Ni-hard paddles - Type 317-LM stainless steel shaft
Fly Ash Silo	2	- 32 ft. diameter 47 ft. high - Carbon steel - 8 Hr. storage capacity - 8 Rotary feeders/silo
Fly Ash Transfer System	2	- Each 160 tons/hr

The Four Corners Waste Processing System incorporates several unique features:

- The Waste Processing System is an elevated plant. The design permits the trucks to drive directly beneath the loading point underneath the processing floor. Blended fly ash and FGD sludge from the process building are transferred directly into the trucks. The high fly ash to sludge ratio noted above makes this material immediately handleable and allows direct truck loading to take place.

- The system is designed to operated on an intermittent basis. Because the loading of the material relies upon the availability of a truck to receive product, it is necessary that the system have the capability of being placed in an idle mode for up to 30 minutes of duration repeatedly during the operating shifts.

SYSTEM DESIGN FEATURES

In 1980, APS commissioned UE&C to perform a study to determine the redundancy, system features, and overall general layout of the Waste Processing System. The objective of this study was to provide a system that met both the redundancy and space limitation requirements of the Four Corners site; and which produces a waste product similar to fly ash alone. This study resulted in the basic configuration of the equipment and structures, general sizing of equipment capacities, and conceptual design of the control philosphy.

DESCRIPTION

FGD System Interface

As an interface with the wet scrubbing system, UE&C provided two thickener underflow tanks, one for each unit with cross connects to allow sparing and common mode capabilities. Each tank is capable of feeding a dedicated processing train as well as the common train. These thickener underflow tanks operate on constant recycle and feed forward from the recycle loop on the thickener underflow tanks. CSI provided a grit strainer and removal system to reduce the amount of foreign material and oversize particles entering into the secondary dewatering system. The objective is to minimize tearing and blinding of the filter cloth and wear on the pumps.

Processing

Each processing train (Figure 2) consists of a slurry overflow feed tank, a rotary drum vacuum filter, a filtercake discharge conveyor, and a pug mill mixer. Each of the slurry overflow feed tanks receives slurry from the recycle loop on the thickener underflow tanks. These tanks then feed forward to the dedicated rotary drum vacuum filter. Each rotary drum vacuum filter is 12' diameter X 22' face for a total of 829 square feet of filtering area. The filter vats operate on constant overflow to the slurry overflow feed tanks.

Each filter discharges filtercake to its dedicated pug mill mixer. The pug mill mixers are rated at 256 tons per hour of throughput capacity. Each mixer contains 44" diameter paddles and is driven by a 150 horsepower motor.

The Ash Transfer System draws ash from the existing fly ash silos, located adjacent to new CSI Waste Processing System. The Ash Transfer System has a rated capacity of 160 tons per hour, and can feed from the old silos to either of two new 848 ton capacity silos located above the mixing floor in the CSI facility. A diverter system at the inlet of the old silos allows the existing Ash Transfer System from the baghouses to feed ash to either the existing fly ash silos or directly to the new fly ash silos.

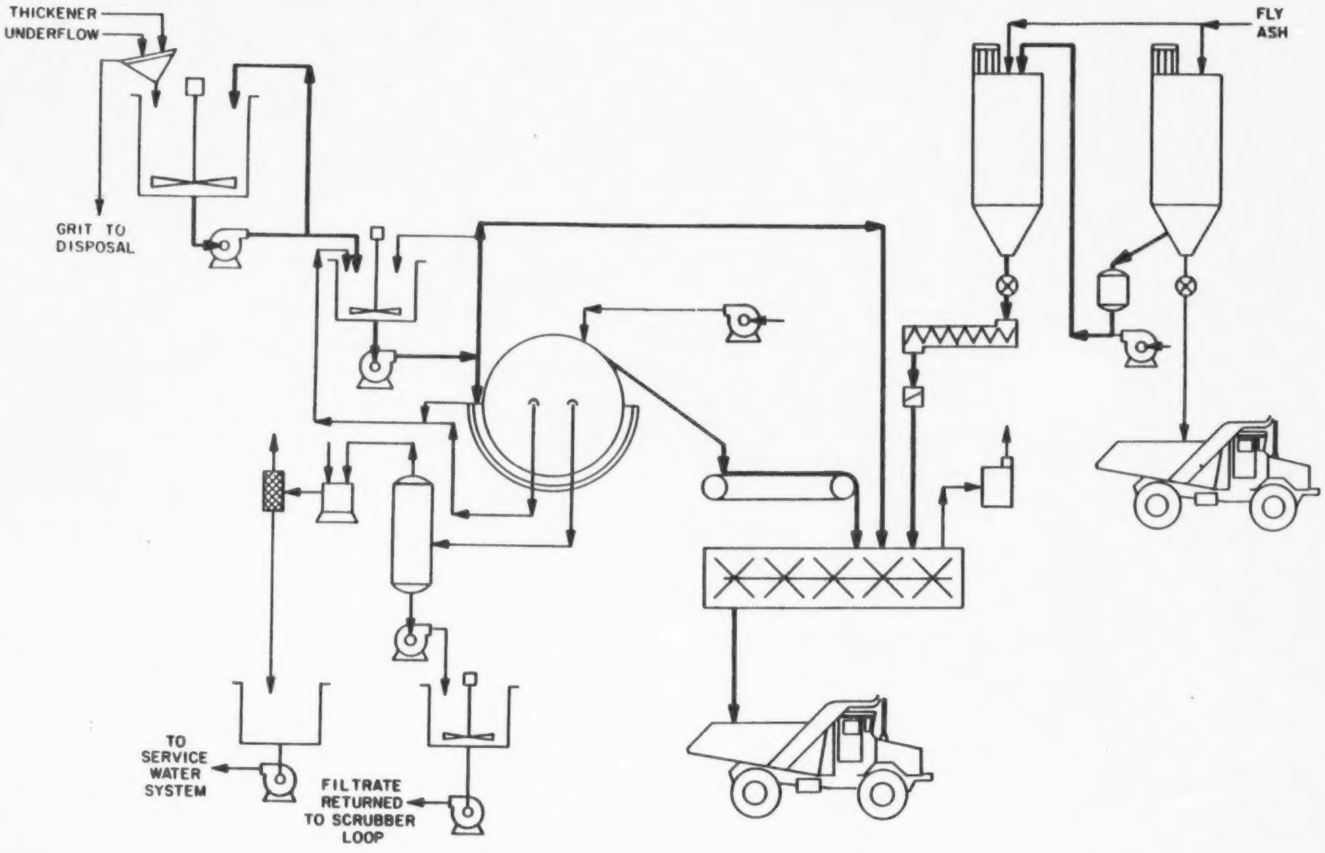

THICKENER UNDERFLOW
GRIT TO DISPOSAL
FLY ASH
TO SERVICE WATER SYSTEM
FILTRATE RETURNED TO SCRUBBER LOOP

Figure 2
ARIZONA PUBLIC SERVICE
FGD WASTE PROCESSING FLOW SHEET

Because the Waste Processing System is designed to operate only two shifts per day, it is sometimes necessary to transfer ash to the old, existing fly ash silos during the eight to eleven hours per day that the Waste Processing System is not operating. Ash is then transferred from the exising silos to the new silos at the beginning of the operating shift. Ash being generated at the baghouses continues to fill the existing silos and the new silos while ash is being drawn from the old silos at the beginning of each operating shift. After approximately five hours of operation the existing silos have been emptied and all ash from the remaining first operating shift and second operating shift flow to the new silos.

Complete redundancy in the ability to feed fly ash is provided by the 16 rotary feeders and 8 screw conveyors that provide ash from the two new fly ash silos to the three pug mill mixers. The fly ash silos are flat bottom, rather than mass flow bin bottom in design and, therefore, include air slides in the silos for fluidization of the ash.

Direct Truck Loading

As originally designed, the blended ash and FGD sludge from the pug mill mixers was discharged to a reversing conveyor located beneath the mixers. This reversing conveyor was a double drive, clutch and brake device with a 20 horsepower motor at each end. The reversing conveyor allowed the blended material from the pug mill mixers to be fed to a truck located in either of two truck loading bays below the processing floor of the system (Figure 3).

An emergency stacker was designed to be engaged to the discharge system by means of a shuttle conveyor connecting in reversing conveyor with the emergency stacker. The emergency stockpile area has the capacity of approximately 2500 cubic yards, or 5.5 hours of blended product storage, at maximum load conditions.

Programmable Controls

This Waste Processing System is fully automatic. The controls are provided by means of a programmable controller, backed up by its own on-line complete spare (Figure 4). A second backup is a hard wire control system to allow manual operation.

The process controls allow the system to be operated in one of five modes:

- ash plus water;
- ash plus slurry;
- ash plus process liquor;
- ash plus filtercake; or,
- ash plus filtercake, plus slurry.

Reproduced from
best available copy.

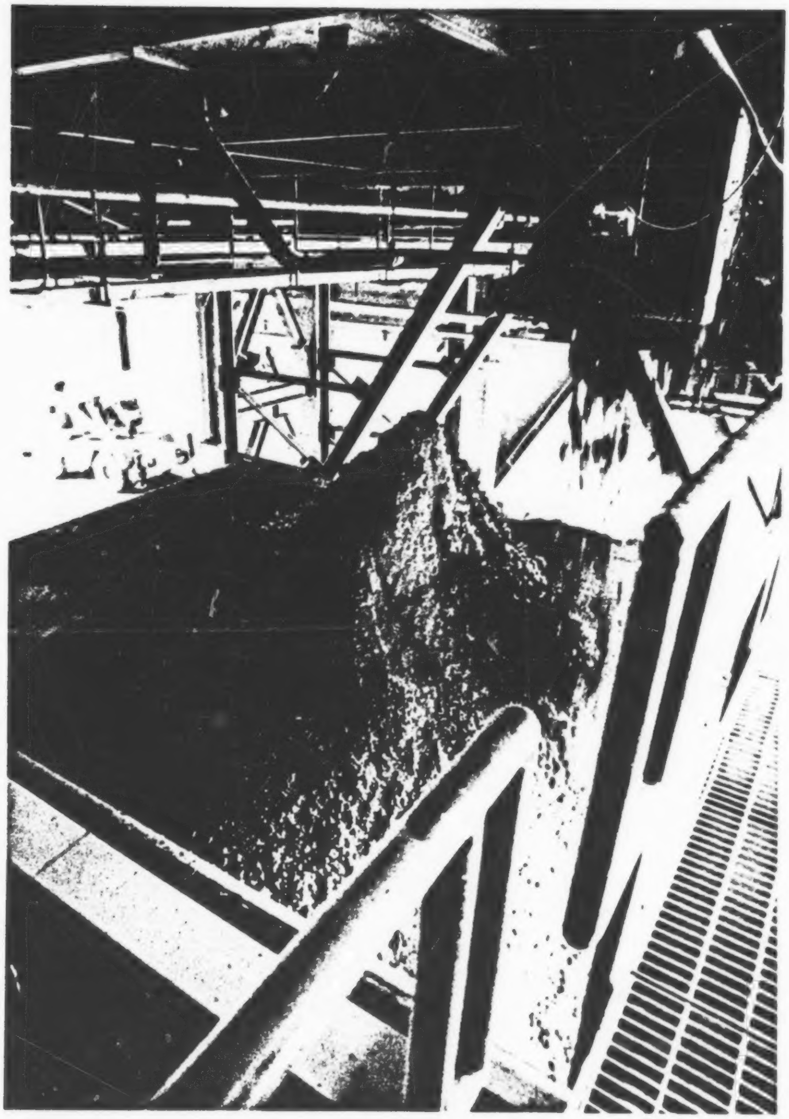


Figure 3 Blended ash and FGD sludge are loaded directly into trucks for disposal in the adjacent mine.

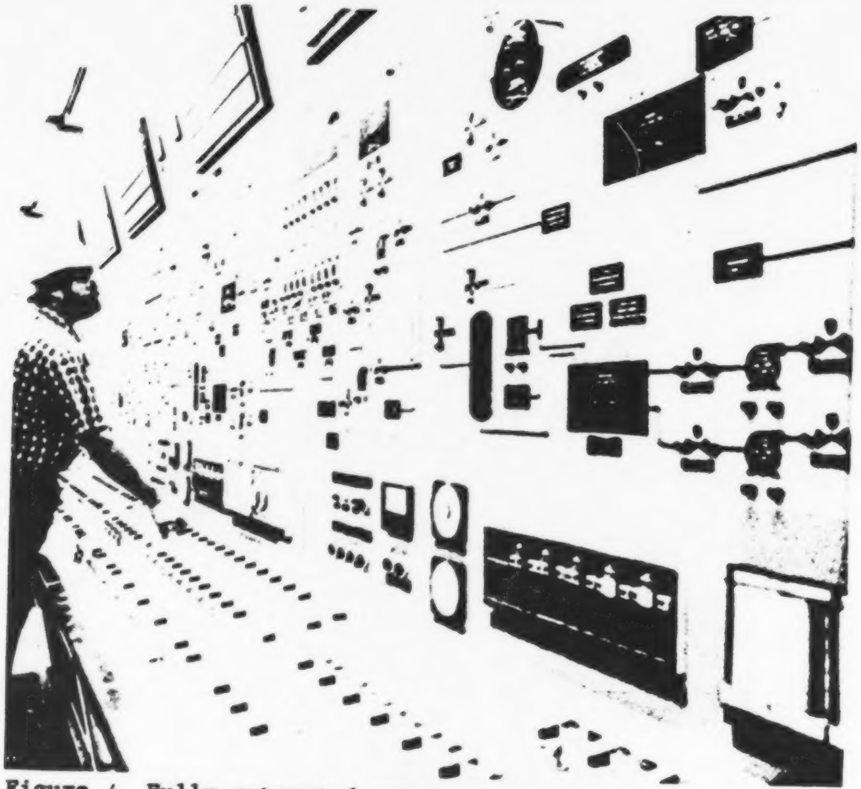


Figure 4 Fully automated programmable controllers allow use of three operations personnel per shift and permit intermittent operation.

The system also incorporates single button features for start, stop and hold modes of operation. The hold feature is designed to accommodate the fact that the discharge of the system is dependent upon the availability of a truck in the receiving bay. The system must have the ability to go into a hold mode for up to thirty minutes of operation during the working shifts. The hold feature is designed in such a way that it may be initiated by either the truck loading operator or the main control panel operator.

The automated controls allow operation of the Waste Processing System by a total of three persons per shift. The operating shift consists of one control panel operator, one roving operator on the process floor, and one truck station loading operator.

Building Design

The vertical design results in a 21' high truck bay. The processing floor and fly ash feeding equipment take the height of the processing building to 58' above grade. The fly ash silos, rotary feeder and screw conveyor area bring the total height of the process facility and silos to 113' above grade, excluding the dust collectors on the fly ash silo (Figure 5).

Reproduced from
best available copy.

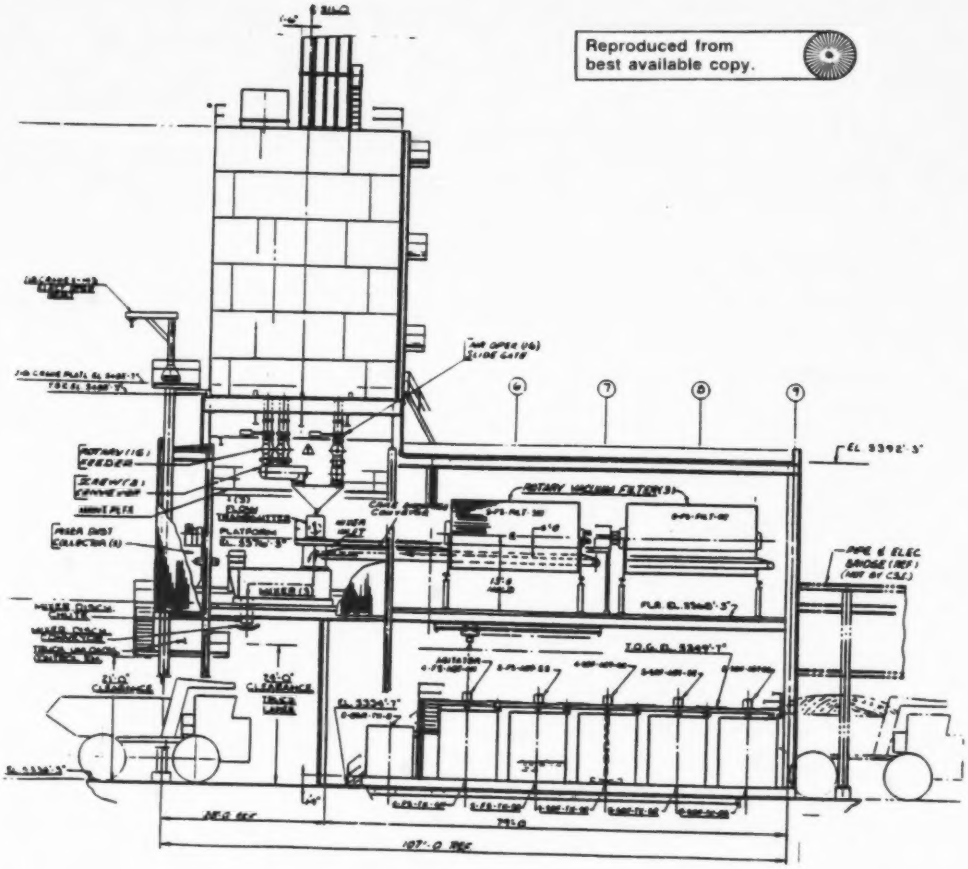

AIR OPER (16) SLIDE GATE
6
7
8
9
EL 5392'-3"
ROTARY (16) FEEDER
SCREW (8) CONVEYOR
PLATFORM
FIBER DUST COLLECTOR (3)
MIXER (3)
MIXER DISCH. CHUTE
ROTARY VACUUM FILTER (3)
13'-6
PIPE & ELEC BRIDGE (REF) (NOT BY C.S.I.)
AGITATOR
21'-0" CLEARANCE
24'-0" CLEARANCE TRUCK LANE
79'-0
107'-0 REF

Figure 5 Elevation drawing shows unique vertical design that allows direct truck loading.

The heating, ventilating and air conditioning (HVAC) system maintains a positive pressure in the process building of 0.125" WC. A monorail and hoist system above all major pieces of equipment within the processing building allows the removal of equipment and its transfer to the area of the maintenance room on the process floor. If it is necessary to lower a piece of equipment to grade for additional service, a large access door and extended monorail with hoist have been provided for this purpose. Jib cranes have also been provided on the roof of the fly ash silos for equipment removal in the ash transfer and the baghouse area.

While minimizing the amount of surface area required for the Waste Processing plant and ash transfer buildings, the vertical design of the system did require 650 tons of structural steel. This compares with approximately 150 tons of structural steel that would have been required for a similar capacity facility at grade.

INITIAL OPERATION

The Waste Processing System began operation in April, 1984. The initial months of operation consisted of blending ash with water. FGD sludge became available for processing in October, 1984. During the first year of operation, the Waste Processing System produced approximately 1,200,000 tons of ash blended with slurry, filtercake, or water. Except for some initially damp material at the start-up of each operation, the final product produced has been comparable to fly ash in handling characteristics.

The start-up of the system was accomplished smoothly and with a minimum of problems. The areas of concern that did occur were primarily in the ash transfer rate and in debugging of the programmable controller. Both of these situations have been resolved and the system operates at design capacity and in the modes for which it was designed to operate. An ongoing evaluation of the type of filter cloth to be used in the operation is being conducted. Currently, 3 different types of filter cloth are being utilized on each of the rotary drum vacuum filters as a test of optimizing secondary dewatering. Selection of the optimum cloth will occur within the next several months.

The reversing conveyor was an area of some concern to APS. This piece of equipment was the only, non-redundant piece of equipment within the plant. In early 1985, APS elected to shorten the reversing conveyor so that it receives blended waste only from the center, common spare processing train. Chutes were added at the discharge of the other two mixers, allowing blended material to be fed to the trucks without the use of the reversing conveyor. In both configurations, the Waste Processing System has not had a negative impact on power generation activities at Four Corners Units 4 or 5.

In addition to the start-up services provided by CSI, Arizona Public Service contracted with CSI for a technical advisor during the first year of operation. It is believed that this advisory service has allowed a smooth and orderly initial operation of the system and has contributed significantly to the better understanding by the APS operators of the functions of the waste processing system.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

CORROSION RESISTANCE COMPARISON OF DUPLEX STAINLESS STEEL WITH TYPE 317LM IN SIMULATED FLUE GAS SCRUBBER ENVIRONMENTS

by: Ronald E. Van Hoose
John B. Guernsey
Eastern Stainless Steel Company
Baltimore, Md. 21203

ABSTRACT

Type 317LM austenitic stainless steel (modified 317L containing 4% minimum molybdenum) is probably the most widely used high alloy material for construction of Flue Gas Desulfurization (F.G.D.) equipment. This is because its corrosion resistance is well suited to handle the corrosive environment produced by F.G.D. systems, with cost far lower than for other suitable alloys. The work discussed here indicates that the austenitic-ferrite duplex stainless steel type 2205 may be equal in corrosion resistance and superior in other regards to type 317LM in F.G.D. applications.

The corrosion resistance of type 2205 and type 317LM were compared in simulated F.G.D. environments to evaluate type 2205 as a construction material for F.G.D. systems. Crevice corrosion, pitting corrosion and general corrosion tests were performed on both types in several different solutions consisting of high chloride concentrations (Na and Ca base), with varying pH levels and high temperatures. Tests to determine critical pitting temperatures of both alloys were also conducted. The results of these tests indicate that the corrosion resistance of type 2205 is comparable to that of 317LM.

The significantly higher strength of type 2205, lower alloy costs and comparable corrosion resistance to simulated F.G.D. environments make it a very good candidate for F.G.D. construction material.

EXPERIMENTAL PROCEDURES

GENERAL PROCEDURES

The chemical composition and mechanical properties of the stainless steel grades tested are given in Table 1. All the steel test coupons for simulated scrubber environment tests were 25 by 19mm (1 by 0.75 inch) in size with a 6.35mm (1/4 inch) hole in the center, polished through 600 grit paper. The crevices were applied by specially prepared high density polyethylene serrated washers. The washers were 12.7mm (0.5 inch) in outside diameter with an inside diameter of 7.1mm (0.28 inch). Grooves in the washer formed 12 separate crevice contacts per washer so that with a washer on each side of the coupon a total of 24 crevices were formed per specimen. A serrated washer is shown in Figure 1. The washers were secured by PVC nuts and bolts torqued to 0.6N·m (15 inch-lb).

Simulated Flue Gas Environments

Test specimens were run in a $CaCl_2$ solution containing 30,000 mg/l Cl at initial pH3 under continuous bubbling of SO_2 and O_2 through the solution. Six hundred twenty five mg of $CaSo_4$ and 625 mg of $CaCo_3$ per liter solution was also in the initial solution, to simulate the slurry inside a scrubber. A test period of two days at 60 C (140 F) was used. The final pH of the test solution was less than 1 due to the introduction of sulfur dioxide and precipitation of calcium sulfite. Test specimens were examined for corrosion under the serrated washers, and depth of all observed corroded areas was measured by means of an optical microscope. The test environment described above was recently employed to evaluate the corrosion resistance in a simulated scrubber environment of several stainless steels and nickel base alloys in both the welded and unwelded condition.(1),(2)

A second simulated scrubber environment was used which consisted of a solution containing 2000 mg/l $NaHSO_3$ and 30,000 mg/l Cl^-(as NaCl) saturated with oxygen by continuously bubbling the gas through the solution. The same kind of crevice test coupons were tested for one week at 80 C (180 F). This test was recently employed to evaluate the crevice corrosion resistance of several stainless steels in a scrubber environment.(3) Duplicate samples were tested at pH levels of 2,3 and 5.3 to simulate three levels of acidity that may occur in scrubbers under normal or upset conditions.

TABLE 1A. COMPOSITION OF STAINLESS STEEL SAMPLES

High Molybdenum
Type 316L - Thickness 0.3750"

C	Mn	P	S	Si	Cr	Ni	Cu	Mo	Co	N
0.012	1.72	0.020	0.028	0.57	17.08	12.08	0.33	2.80	0.22	0.036

Type 317LM - Thickness 0.3125"

C	Mn	P	S	Si	Cr	Ni	Cu	Mo	Co	N
0.022	1.30	0.020	0.018	0.63	18.43	15.11	0.25	4.38	0.22	0.045

Type 904L - Thickness 0.3125"

C	Mn	P	S	Si	Cr	Ni	Cu	Mo	Co	N
0.020	1.64	0.019	0.006	0.41	20.08	24.25	1.35	4.54	0.31	0.021

Type ES2205 - Thickness 0.1875"

C	Mn	P	S	Si	Cr	Ni	Cu	Mo	Co	N
0.018	0.48	0.026	0.015	0.40	22.62	5.51	0.20	3.38	0.11	0.141

TABLE 1B. MECHANICAL PROPERTIES OF STAINLESS STEEL SAMPLES (Anneal/Typical)

Type	Tensile Strength p.s.i	Yield Strength (offset: 0.2%) p.s.i.	Elongation % in 2 inches	Hardness
316L	86,000	44,000	53%	R_B 85
317LM	85,000	40,000	48%	R_B 84
904L	85,000	43,000	45%	R_B 80
ES2205	120,000	82,000	30%	R_C 20

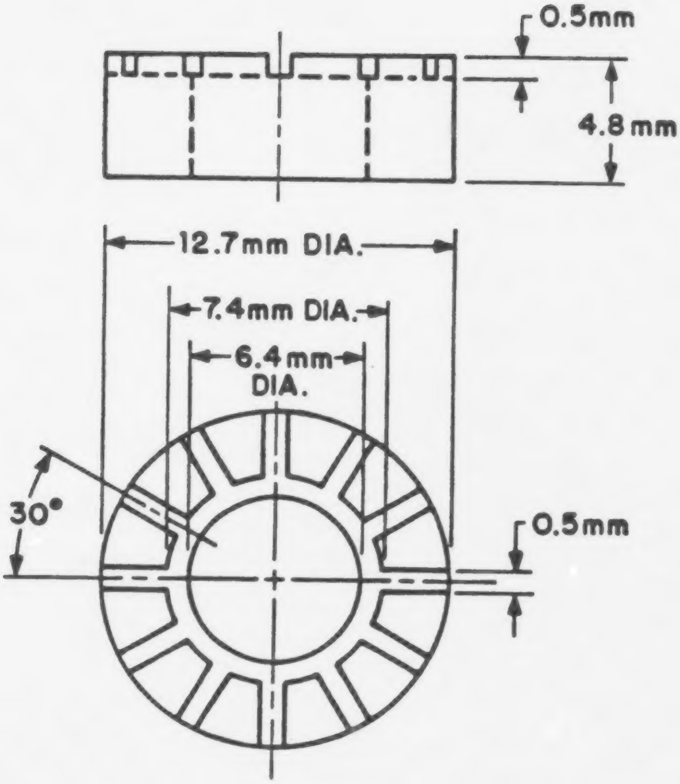

0.5mm
4.8mm
12.7mm DIA.
7.4mm DIA.
6.4mm DIA.
30°
0.5mm

Figure 1 - Design of a Serrated Washer

Pitting Corrosion Reistance

Resistance to pitting corrosion was determined by conducting anodic polarization experiments at a scan rate of 0.6V/hour, from-0.8V versus SCE until pitting occurred. A deoxygenated aqueous test solution containing 0.01M H_2SO_4 + 15,000 mg/l Cl^- (as NaCl) at 50 C (122 F) was used. The test samples used were machined to 6.5 X 6.5 X 3mm (0.25 X 0.25 X 0.125"), mounted in epoxy and polished through 600 grit paper. To prevent corrosion of the interface, the sample/mount interface was covered with Glyptal. This solution was used because chlorides and sulfuric acid are the major constituents in scrubbers that contribute to corrosion attack.

RESULTS

Calcium-Based Solution

The results of the two day test in the solution containing 30,000 mg/l Cl^- (as CaCl), 625 mg/l $CaSO_4$, 625 mg/l $CaCO_3$ with continuous bubbling of sulfur dioxide and oxygen at 60 C (140 F) are detailed in Table 2. This proved to be a mild test. Only one of two 2205 duplex stainless steel samples showed some attack. Only one crevice site was attacked (out of 24 sites available) with a maximum pit penetration of 26 microns. The degree of pit penetration in 2205 is comparable to that of the other austenitic stainless steel samples. The corrosion rates of type 317LM, type 904L and and 2205 were 1-2 mg/clm^2/day, while the corrosion rate of high-molybdenum type 316L was approximately 4 mg/dm^2/day.

Sodium-Based Solution

The results of the seven day test in the solution containing 2,000 mg/l $NaHSO_3$ + 30,000 mg/l Cl^- (as NaCl) saturated with continuous bubbling of oxygen at 80 C (180 F) are shown in Table 3. Crevice corrosion was observed in only one to eight crevice sites (out of 24 sites available) even in this more severe test environment. As shown in the table and in Figure 2, the maximum pit penetration was in the range of 20 to 120 microns for type 317LM, type 904L and type 2205. Type 316L's maximum pit penetration was much higher, especially at a pH level lower than 3. Type 2205 had the lowest pit penetration at the 2pH level of all the samples tested. The corrosion rates as a function of the pH are shown in Figure 3, and again 2205 is as good as any of the alloys tested at low pH levels.

Pitting Corrosion Resistance

The results of the pitting corrosion tests conducted in the 0.01m H_2SO_4 and 25,000 mg/l Cl^- (as NaCl) at 50 C (122 F) are show in Table 4. These anodic polarization tests indicate that the pitting resistance of type 2205 duplex stainless steel is comparable to that of type 317LM and type 904L. The results also show the pitting resistance of type 316L to be far less than for the other three alloys tested.

TABLE 2

Crevice Corrosion Tests in 30,000 ppm Cl^- (As $CaCl_2$),
Initial pH=3, Saturated with $O_2(g)$ and $SO_2(g)$, 2 Days at 60 C (140 F)

Steel Type	Number of Crevice Sites Attacked (out of 24 Total)	Maximum Pit Depth in the Creviced Area (Microns)	Corrosion Rate ($mg/dm^2/day$)
316L	0	0	4.4
	1	66	4.0
317LM	1	7	1.8
	1	41	1.1
904L	1	12	1.1
	4	66	1.4
ES2205	0	0	1.5
	1	26	2.2

TABLE 3

Crevice Corrosion Behavior of Stainless Steels in 30,000 ppm Cl^-, 2,000 ppm $NaHSO_3$ Saturated with Oxygen, 7 Days Test

pH	Steel Type	Number of Crevice Site Attacked (Out of 24 Total)	Maximum Pit Depth in the Creviced Area (Microns)	Corrosion Rate $mg/dm^2/day$
2	316L	4*	528	189
		5*	402	149
	317LM	6	42	0.9
		8	93	1.8
	904L	2	24	0.7
		1	215	83.9
	ES2205	1	36	1.4
		4	42	0.6
3	316L	6*	667	258
		7*	414	151
	317LM	2	46	0.6
		8	38	0.1
	904L	1*	86	45.6
		0	0	0.6
	ES2205	2	38	0.4
		3	48	3.1
5.3	316L	1*	175	17.1
		7*	87	43.9
	317LM	1	13	0.0
		7	34	1.0
	904L	1	19	0.6
		7	61	0.8
	ES2205	2	39	0.6
		8	46	89.2

*Specimens were also subject to general corrosion.

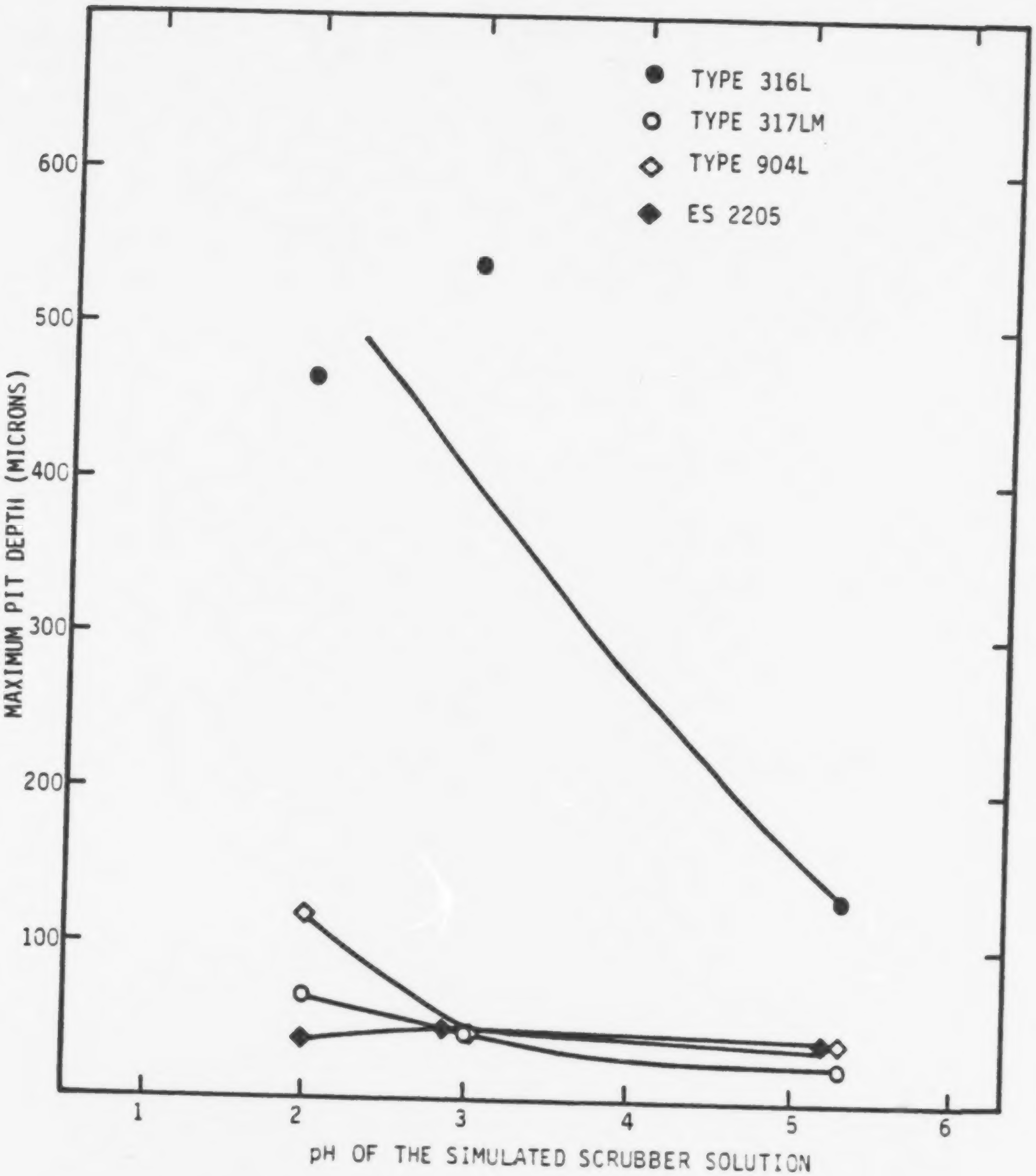


Figure 2 - Maximum Pit Depth as a Function of pH of a Simulated Scrubber Solution Containing 2000 mg/l $NaHSO_3$ and 30,000 mg/l Cl^- Saturated with Oxygen at 80°C (180°F)

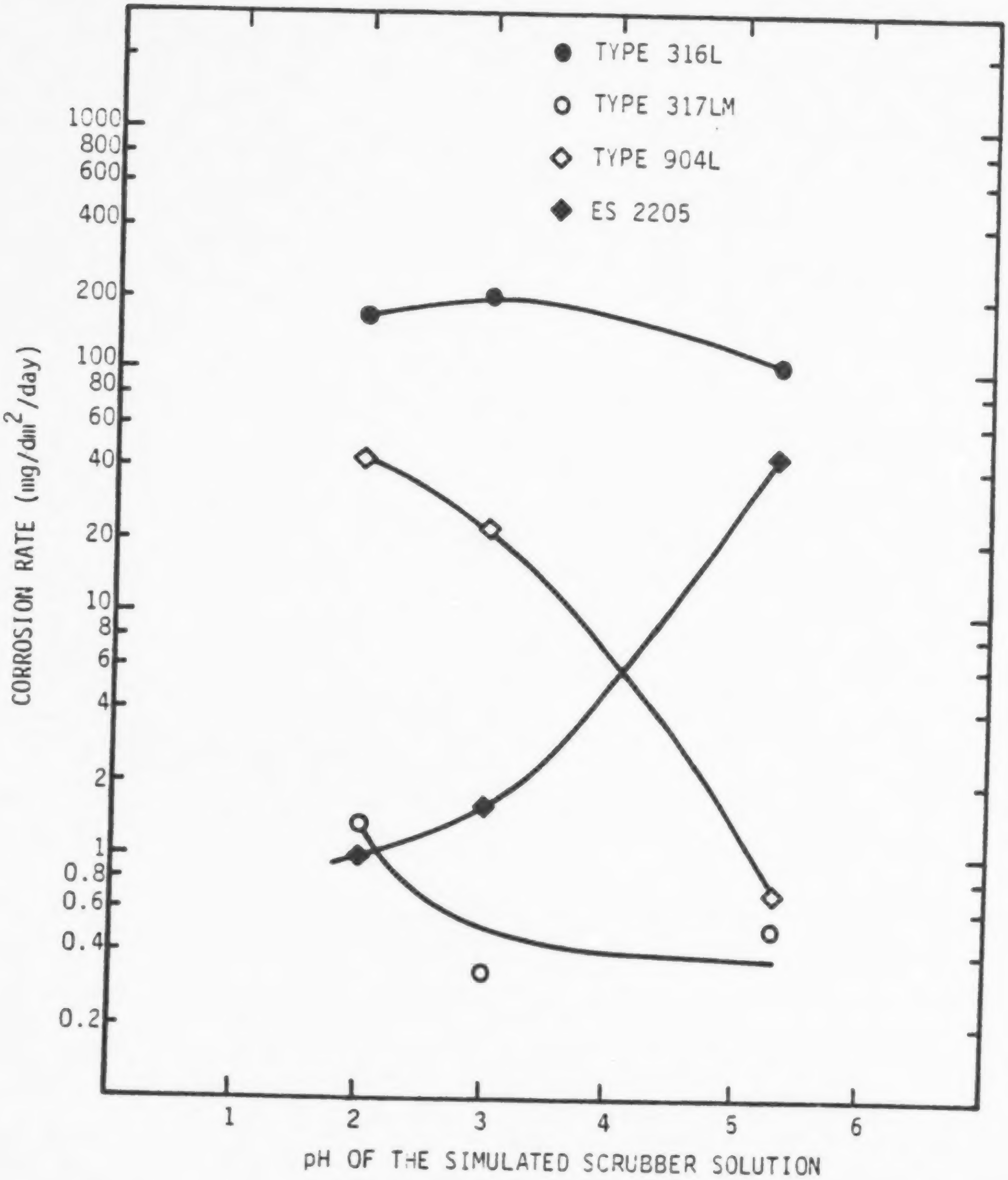


Figure 3 - Corrosion Rate as a Function of pH of a Simulated Scrubber Solution Containing 2000 mg/l $NaHSO_3$ and 30,000 mg/l Cl^- Saturated with Oxygen at 80°C (180°F)

TABLE 4

Pitting Potentials of Stainless Steels (vs. SCE) at 50 C (122 F) Determined at a Scan Rate of 0.6 V/h in a Dearated Solution Containing 25,000 ppm Cl^- (as NaCl) + 0.01M H_2SO_4

Steel Type	Pitting Potential (mv)	Average Pitting Potential (mv)
316L	-40, 80, 65	35
317LM	225, 220, 175	207
904L	120, 240	180
ES2205	230, 170	200

SUMMARY

The results of the crevice corrosion tests conducted in the two different simulated Flue Gas Scrubber solutions indicate that the corrosion resistance of type 2205 duplex stainless steel is comparable to that of type 317LM and type 904L and superior to that of type 316L stainless steel. The results also show type 2205 to have a lower corrosion rate than type 904L at a pH level of 2 and 3.

The results of the pitting corrosion tests also indicate very comparable corrosion rates among types 2205, 317LM and 904L, which are all superior to type 316L.

The comparison of mechanical properties shows the much higher strength level of 2205 over type 317LM, type 316L and 904L. This means that lighter gages (less tonnage required) could be used for construction where design permits.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

REFERENCES

1. E.A. Lizlovs, "Corrosion Resistance of Commercial High Cr-Mo Austenitic Stainless Steels and other Alloys in a Simulated SO_2-Scrubber Environment," Climax Report RP-22-81-02, May 20, 1985.

2. A. Poznansky, "Crevice Corrosion of Welded Stainless Steels and Nickel Base Alloys in a Simulated Scrubber Environment," Climax Report RP-33-82-09, July 15, 1985.

3. E. A. Lizlovs, "Corrosion Resistance of types 316 and 317 Stainless Steels in Simulated Scrubber Environment", Journal of Electrochemical Society; Vol. 124, Nov. 12, 1977.

CORROSION RESISTANCE OF A Cr-Ni-Mo-Cu-Fe ALLOY IN SCRUBBER ENVIRONMENTS AND OTHER CHEMICALS

by: Terry A. DeBold, Douglas G. Frick, Gerald A. Bauer
Carpenter Technology Corporation
P.O. Box 662
Reading, PA 19603-0662

ABSTRACT

A new austenitic stainless steel with optimized alloy content has been developed which provides a high level of resistance to scrubber environments and other chemicals. Laboratory data for several forms of attack including pitting, crevice, intergranular, and general corrosion are provided along with mechanical properties. Spool test data from several scrubber locations are included.

INTRODUCTION

Increasing chloride levels in SO_2 scrubbers have led to the use of more highly alloyed materials with a corresponding increase in cost. Materials become more attractive as their corrosion resistance is maximized with minimum costly alloy additions. This paper presents data on one such alloy, 20Mo-6[1] stainless (U.S. Patent 4,201,575), and compares it to other materials which are useful in all but the most severe scrubber environments.

Carpenter 20Mo-6 stainless is an austenitic stainless steel which is resistant to pitting, crevice and stress corrosion in hot, low-pH chloride environments. It is also resistant to oxidizing media such as hot nitric acid. Carpenter 20Mo-6 stainless complements 20Cb-3[1] stainless, but does not replace it. The new alloy is designed for applications where better pitting and crevice corrosion resistance are required than 20Cb-3 stainless offers.

Laboratory and field test data are provided to characterize the corrosion properties of 20Mo-6 stainless. Comparison data for Hastelloy[2] alloys G and G-3 and Alloy 904L are included when available. Nominal compositions for these materials are in Table 1.

TABLE 1. NOMINAL COMPOSITIONS*

	C	Mn	Si	Cr	Ni	Mo	Cu	Cb+Ta	Other
20Mo-6 stainless	0.02	0.4	0.35	23.7	36.5	5.7	3.2	--	--
Hastelloy alloy G	0.03	1.3	0.35	22.2	44	6.5	2.0	2.12	.50 W
Hastelloy alloy G-3	.015 max	0.8	0.4	22.2	44	7.0	1.9	0.3	1.5 W max. 5.0 Co max.
Alloy 904L	0.02	1.7	0.35	20.0	25.0	4.5	1.5	--	--

*Weight percent, balance Fe.

EXPERIMENTAL PROCEDURE

Most of the work was performed on mill annealed sheet or plate. Laboratory samples were machine ground to 120 grit and unless otherwise noted were finished to 38.1 mm x 12.7 mm x 3.2 mm (1-1/2" x 1/2" x 1/8"). Pitting and crevice samples were 25.4 mm x 50.8 mm (1" x 2") as stipulated by ASTM G 48.(1) U-bends were 98.4 mm x 9.5 mm x 3.2 mm (3-7/8" x 3/8" x 1/8"). They were bent to a 25.4 mm (1") diameter and were insulated from the stressing bolt with Teflon[1]. Field data were obtained using spool test racks(2) and samples with a mill finish.

---

[1] Registered trademark, Carpenter Technology Corporation

[2] Registered trademark, Cabot Corporation

Tube samples were cut from GTAW tubing and finished to 120 grit. For consistency, GTAW was also used for welded strip samples. These coupons contained a 25.4 mm (1") length of weld and were 31.8 mm (1-1/4") wide.

GENERAL, INTERGRANULAR, AND STRESS CORROSION

GENERAL CORROSION RESISTANCE

Carpenter 20Mo-6 stainless, Hastelloy alloy G, and Alloy 904L were exposed to various concentrations of reagent grade sulfuric acid at 80°C (176°F) for three 48-hour periods. Corrosion rates are listed in Table 2. Carpenter 20Mo-6 stainless and Hastelloy alloy G were generally similar and had good resistance except in more highly concentrated acid. Alloy 904L was inferior to the other alloys except in 10% acid. Resistance of all three grades may improve if exposed to commercial sulfuric acid. Many contaminants in commercial sulfuric acid can decrease corrosion. Corrosion rates in sulfuric acid can be lower in the presence of iron, copper, and chromium ions, usually present in pickling and plating solutions. Characterization of the stress corrosion cracking resistance of 20Mo-6 stainless and comparison materials in sulfuric acid is in progress.

TABLE 2. RESISTANCE IN SULFURIC ACID AT 80°C (176°F)

Concentration (w/o)	20 Mo-6 stainless*		Hastelloy alloy G*		Alloy 904L*	
	mm/y	mpy	mm/y	mpy	mm/y	mpy
10	0.03	1	0.08	3	0.03	1
40	0.18	7	0.20	8	6.60	260
45	0.20	8	0.23	9	1.90	75
50	0.20	8	0.23	9	1.22	48
90	0.96	38	1.35	53	3.88	153
96	1.35	53	0.43	17	3.30	130

*Average of three 48-hr. exposures of duplicate samples.

Annealed samples of all four alloys were tested in wet-process phosphoric acid at 100°C (212°F) for three 48-hour exposures. The composition of the acid is listed with corrosion rates in Table 3. All materials were resistant, although Alloy 904L had an average corrosion rate about twice that of the other grades.

---

[1] Registered trademark, E. I. duPont de Nemours & Co.

TABLE 3. RESISTANCE IN WET PROCESS PHOSPHORIC ACID* AT 100°C (212°F)

Alloy	Average of Three 48-hr. Exposures mm/y	mpy	Appearance
20Mo-6 stainless	0.08	3	Light general attack
Hastelloy alloy G	0.10	4	Light general attack
Hastelloy alloy G-3	0.08	3	Light general attack
Alloy 904L	0.18	7	Light general attack

*68.9 w/o H_3PO_4, 4.15 w/o H_2SO_4, 0.54 w/o F^-, 1.85 w/o Fe, 2000 ppm Cl^-
Samples activated with zinc rod at test beginning.
Each value represents the average of two to four samples.

Test spool assemblies of Carpenter 20Mo-6 stainless, Hastelloy alloy G, and Alloy 904L were exposed to wet-process superphosphoric acid in a two stage evaporator for 87 days. One spool was tested in the first stage evaporator in boiling acid at 160°C (320°F). Another spool was exposed in the second stage evaporator at 188°C (370°F). The acid contained 70 w/o phosphoric acid, 3 w/o sulfate, and 0.6 w/o fluoride. Results are in Table 4. Based on weight loss, Carpenter 20Mo-6 stainless and Hastelloy alloy G were slightly superior to Alloy 904L in these tests. The crevice attack depth for all alloys in Stage I was less than .25 mm (.010") deep. This test could not be used to rank the alloys because the spool was bent, apparently before complete exposure, resulting in non-uniform geometry of the crevice sites. Small pits formed on some of the alloys in the second stage.

TABLE 4. RESISTANCE OF TEST SPOOLS IN A WET PROCESS SUPERPHOSPHORIC ACID* EVAPORATOR

	Weight Loss in mg/cm^2	
Alloy	First Stage Evaporator 160°C (320°F)	Second Stage Evaporator † 188°C (370°F)
20Mo-6 stainless	34	100
Hastelloy alloy G	35	83
Alloy 904L	55	158

*70 w/o H_3PO_4, 3 w/o SO_4^{-2}, 6000 ppm F^-, 87 day test.
†Residue formed on surface of samples during exposure.
The residue contained 81,000 ppm chloride and 1500 ppm fluoride.

INTERGRANULAR CORROSION

The intergranular corrosion resistance of 20Mo-6 stainless was evaluated for several material conditions using the ferric sulfate-sulfuric acid test (ASTM A262 - Practice B). Results in Table 5 demonstrate that 20Mo-6 stainless has good intergranular corrosion resistance. The corrosion rate of welded strip sensitized at 538°C (1000°F) for 24 hours was above that for annealed material. These data suggest that stress relieving composite articles containing carbon steel be conducted below 538°C (1000°F) for less than 24 hours.

TABLE 5. RESISTANCE OF 20Mo-6 TO FERRIC SULFATE - SULFURIC ACID*

Heat Treatment[†]	Product Form	Corrosion Rate	
		mm/y	mpy
Annealed	Strip	0.20	8
Welded	Strip	0.20	8
Welded + 538°C (1000°F), 24 h.	Strip	0.48	19
Welded	Tube	0.20	8
Welded + Annealed	Tube	0.25	10

*ASTM A262, Practice B.

[†] Autogenous GTAW process used for welded samples.

Table 6 provides comparison data for 20Mo-6 stainless, Hastelloy alloy G, and Alloy 904L exposed to the ferric sulfate-sulfuric acid test following a 677°C (1250°F) sensitization. The difference in rates for the three alloys was due to variation in general and intergranular attack. Alloy 904L suffered the deepest intergranular attack.

TABLE 6. COMPARISON OF THREE ALLOYS IN FERRIC SULFATE - SULFURIC ACID*

Alloy	Corrosion Rate	
	mm/y	mpy
20Mo-6 stainless	0.28	11
Hastelloy alloy G	0.74	29
Alloy 904L	1.12	44

*ASTM A262, Practice B.
Sensitized 677°C (1250°F), 1 hour, air-cooled.

STRESS CORROSION CRACKING

Triplicate U-bend samples of 20Mo-6 stainless tested in 35 w/o boiling magnesium chloride at 125°C (257°F) failed in 864 hours. In the same series of tests, 20Cb-3 stainless failed in 216 hours. In the nearly 20 years of field experience with 20Cb-3 stainless there has never been a documented case of chloride stress corrosion cracking. Therefore, the longer failure times for 20Mo-6 stainless indicate excellent resistance to chloride stress corrosion cracking. More concentrated magnesium chloride causes more rapid failures which can be misleading in view of the absence of field failures of 20Cb-3 stainless.

PITTING AND CREVICE CORROSION RESISTANCE

LABORATORY DATA

Pitting and crevice corrosion resistance was evaluated in 6 w/o $FeCl_3$ (10 w/o $FeCl_3.6H_2O$) for three days according to ASTM G 48. All four alloys were tested at room temperature and 50°C (122°F). The results are presented in Table 7. In these tests, 20Mo-6 stainless and Hastelloy alloys G and G-3 exhibited good resistance while higher weight losses were obtained for alloy 904L.

TABLE 7. RESISTANCE TO PITTING AND CREVICE CORROSION IN 6 w/o $FeCl_3$*

Alloy	Weight Loss in mg/cm^2† Room Temperature Pitting	Room Temperature Crevice	50°C (122°F) Pitting
20Mo-6 stainless	<0.01	0.5	1.1
Hastelloy alloy G	<0.01	0.06	<0.01
Hastelloy alloy G-3	<0.01	<0.01	NT ¶
Alloy 904L	0.01	4.7	46.3

*ASTM G48.
†Each value represents the average of two to 24 samples.
¶Not tested.

Pitting temperatures were obtained for all materials using the solution and specimen geometry from G 48. Samples were weighed and exposed for 24 hours to the $FeCl_3$ solution beginning at room temperature. Coupons which were unattacked were immersed in fresh $FeCl_3$ solution at 2.5°C (4.5°F) higher than the previous immersion. After 24 hours they were removed, weighed, and examined visually for attack. This process was repeated until attack was noted and the lowest temperature at which pitting occurred was recorded. Results are in Table 8. In these tests, 20Mo-6 stainless was better than Alloy 904L but inferior to Hastelloy alloys G and G-3.

TABLE 8. PITTING TEMPERATURES IN 6 w/o FERRIC CHLORIDE

Alloy	Pitting Temperature °C	Pitting Temperature °F
20Mo-6 stainless	54.0	129.2
Hastelloy alloy G	62.5	144.5
Hastelloy alloy G-3	65.0	149.0
Alloy 904L	44.5	112.1

Additional pitting temperature tests were conducted in an oxidizing NaCl-HCl environment described by Manning.(3) The solution contained 4 w/o NaCl + 0.1 w/o $Fe_2(SO_4)_3$ + 0.01M HCl. All four alloys were tested in two conditions: mill annealed and welded (GTAW with filler). Carpenter 20Mo-6 stainless and Alloy 904L were welded with Alloy 625 (ERNiCrMo-3). Hastelloy alloys G and G-3 were welded with Hastelloy alloy G-3. Test specimen dimensions were described previously. Duplicate samples of each alloy in each condition were tested according to the test technique described for the ferric chloride pitting temperature tests. The results of the tests are shown in Table 9. Carpenter 20Mo-6 stainless, Hastelloy alloy G, and Hastelloy alloy G-3 were highly resistant and all three were superior to Alloy 904L.

TABLE 9. RESISTANCE TO PITTING IN AN OXIDIZING NaCl - HCl ENVIRONMENT*

Alloy	Condition	Pitting Temperature † °C	Pitting Temperature † °F
20Mo-6 stainless	Mill Annealed	62.5	144.5
Hastelloy alloy G	Mill Annealed	67.5	153.5
Hastelloy alloy G-3	Mill Annealed	67.5	153.5
Alloy 904L	Mill Annealed	47.5	117.5
20Mo-6 stainless	Welded	62.5 ¶	144.5
Hastelloy alloy G	Welded	60.0 §	140.0
Hastelloy alloy G-3	Welded	67.5 #	153.5
Alloy 904L	Welded	47.5 Φ	117.5

---

*4 w/o NaCl + 0.1 w/o $Fe_2(SO_4)_3$ + 0.01M HCl.
†Lowest temperature at which pitting occurred.
¶Both parent metal and weld pitted.
§Pits in heat affected zone.
#Weld pitted.
ΦPits in parent metal.

Samples without crevices were exposed to an acid-chloride mixture containing sulfuric acid, hydrochloric acid, cupric chloride, and ferric chloride.(4) Tests were conducted at 70°C (158°F) for 48 hours. Results in Table 10 show that 20Mo-6 stainless was approximately equal to Hastelloy alloy G and superior to Alloy 904L.

TABLE 10. RESISTANCE IN A LOW pH-CHLORIDE ENVIRONMENT* AT 70°C (158°F)

Alloy	Weight Loss in 48 hr. - mg/cm^2	Appearance
20Mo-6 stainless	22/27	Random Pitting
Hastelloy alloy G	30/31	Random Pitting
Alloy 904L	38/39	Random Pitting

---

*7 v/o H_2SO_4, 3 v/o HCl, 1 w/o $CuCl_2$, 1 w/o $FeCl_3$, balance H_2O.

Coupons of the four alloys were tested in three simulated scrubber solutions. Initial solution compositions along with the corrosion resistance of the alloys are presented in Table 11. In solutions A and B with higher halogen levels, the corrosion rates of 20Mo-6 stainless, Hastelloy alloy G, and Hastelloy alloy G-3 are nearly equivalent while the rates for alloy 904L are considerably higher. Samples with weight losses experienced general attack or etching. All four alloys were unattacked in the most dilute solution.

TABLE 11. RESISTANCE OF ANNEALED SPECIMENS IN SIMULATED SCRUBBER SOLUTIONS*

Alloy/Sol'n. Component	Solution A		Solution B		Solution C	
	mm/y	mpy	mm/y	mpy	mm/y	mpy
20Mo-6 stainless	0.85	35.0	<0.01	<0.1	<0.01	<0.1
Hastelloy alloy G	0.84	33.1	<0.01	<0.1	<0.01	<0.1
Hastelloy alloy G-3	0.75	29.7	<0.01	<0.1	<0.01	<0.1
Alloy 904L	1.42	55.9	0.63	24.7	<0.01	<0.1
Cl^-	136,000		47,800		13,900	
F^-	1,160		382		160	
$SO4^=$	45,300		24,200		13,400	
Na	80,400		26,900		7,760	
pH	1		1		1	
°C	54		54		54	

---

*336 - hour exposure.
Chemical analyses performed before corrosion testing.
Solution compositions are in micrograms per milliliter.

FIELD DATA

Two test spools were tested in a wet lime scrubber system in an eastern United States power plant for 85 days. One spool was exposed at the stack entrance in the mixing zone at 66 to 71°C (150 to 160°F). The second spool was located at an absorber outlet in wet, saturated flue gas at 49 to 54°C (120 to 130°F). This spool was heavily coated with fly ash containing chloride and fluoride. The pH of the outlet ducting flue gas condensate was 1 to 3. The scrubbers contained about 1500 ppm chloride.

The samples were cleaned ultrasonically in detergent solution, weighed, and the pit and crevice depths were measured metallographically. Samples with heavy fly ash were scraped to remove the deposit. Results for duplicate samples of 20Mo-6 stainless, Hastelloy alloy G, and Alloy 904L are in Table 12. At the stack entrance, Hastelloy alloy G was slightly superior to 20Mo-6 stainless and both alloys were vastly superior to Alloy 904L. Attack was much more severe at the absorber outlet where 20Mo-6 stainless was superior to both Hastelloy alloy G and Alloy 904L.

TABLE 12. RESISTANCE OF TEST SPOOLS IN WET LIME SCRUBBER SYSTEM*

	Stack Entrance 66/71°C (150/160°F)			Absorber Outlet 49/54°C (120/130°F)†		
	Max Pit/ Crevice Attack			Max Pit/ Crevice Attack		
Alloy	mm	mils	Average mg/cm^2	mm	mils	Average mg/cm^2
20Mo-6 stainless	0.36	14	1	0.56	22	7
Hastelloy alloy G	0.10	4	1	1.68¶	66¶	8
Alloy 904L	0.91§	36§	5	0.91§	36§	29

*About 1500 ppm Cl$^-$ in scrubbers; pH 1 to 3 in outlet ducting condensate; flue gas velocity about 70 to 90 ft./sec.; 85 day exposure of duplicate samples.

†Fly ash covered spool at test completion. The fly ash contained 25 w/o sulfate, 11,500 ppm fluoride, and 2700 ppm chloride.

¶Perforated

§Significant general corrosion along with pitting and crevice attack.

Spool tests were also conducted for 197 days in a closed loop limestone scrubber system at the R. D. Morrow, Sr., generating plant. Samples were placed at several locations: the absorber, absorber inlet and outlet, bypass duct, and stack entrance. The absorber slurry contained 20,000 ppm chloride at pH 5 and 52°C (125°F). The average temperature at the stack entrance was 104°C (219°F). A sludge deposit on the stack breeching contained 108,000 ppm fluoride and 8300 ppm chloride. Results in Table 13 have been extracted in part from other work.(5,6) Carpenter 20Mo-6 stainless was equal or superior to Hastelloy alloy G in four of the five locations. Alloy 904L suffered the least attack of the three alloys in the stack entrance.

TABLE 13. CORROSION TEST SPOOL RESULTS IN A CLOSED LOOP LIMESTONE SYSTEM[*]

	Maximum Pit/Crevice Attack					
	20Mo-6 stainless		Hastelloy alloy G		Alloy 904L	
Spool Location	mm	mils	mm	mils	mm	mils
Absorber Inlet	0.01	1	0.13	5	0.03	1
Absorber	Nil	Nil	0.08	3	Nil	Nil
Absorber Outlet	0.25	10	0.05	2	0.33	13
Bypass Duct	Nil	Nil	0.03	1	0.03	1
Stack Entrance	0.33	13	0.53	21	0.15	6

[*] Exposure duration 197 days; Absorber Slurry 20,000 ppm Cl^-, pH = 5, 52°C (125°F); Stack Entrance average temperature 104°C (219°F).

TABLE 14. RESISTANCE OF TEST SPOOLS IN THE DALLMAN FGD SYSTEM[*]

Spool Location	pH	Max. Temp. °C	Max. Cl ppm	20Mo-6 stainless Max Local. Attack, mm	20Mo-6 stainless Rate mm/y	Hastelloy alloy G Max Local. Attack, mm	Hastelloy alloy G Rate mm/y	Alloy 904L Max Local. Attack, mm	Alloy 904L Rate mm/y
Quencher Sump (Submerged; 1.8m level)	4.4	60	7000	0	0.0	< .02	0.0	0	0.0
Quencher Sump (Submerged; 3.4m level)	4.4	60	7000	0	0.0	< .03	0.002	0	0.0
Quencher Spray Header, above Slurry	4.4	60	100	0.15[¶]	0.02	0.10[†]/0.61[¶]	0.02	0.25[¶]	0.03
Absorber, Spray Area	6.2	60	100	0	0.0	0.03	0.0	0	0.0
Outlet Duct	2-4[§]/1.5[#]	55[§]/168[#]	100[§]/82000[#]	0.36[Φ]	0.01	0.99[†]	0.01	0.53[†]	0.01

[*] Slurry: 7000 ppm dissolved Cl^-.
Quencher, Inlet Duct, Absorber, Outlet Ducting: Deposits contained 3000 to 4000 ppm Cl^- and 800 to 1900 ppm F^-.
Inlet and Outlet Duct Areas: pH 1-4 and 110-149°C (230-300°F).
Duration: 6 months, 39 days on bypass.

[†] Profuse
[¶] Sparse
[§] Operation
[#] Bypass
[Φ] Intergranular etch

Paul and Ross reported results for spools exposed to the V. Y. Dallman flue gas desulfurization (FGD) system in Springfield, IL.(7) Spool locations included the quencher sump, the quencher spray header above the slurry, the absorber, and the outlet duct. Results for 20Mo-6 stainless, Hastelloy alloy G, and Alloy 904L are presented in Table 14. In these environments, 20Mo-6 stainless was equal or superior to the other two alloys.

MECHANICAL PROPERTIES

Typical room and elevated temperature mechanical properties for 6.4 mm (1/4") plate of 20Mo-6 stainless are shown in Figure 1.

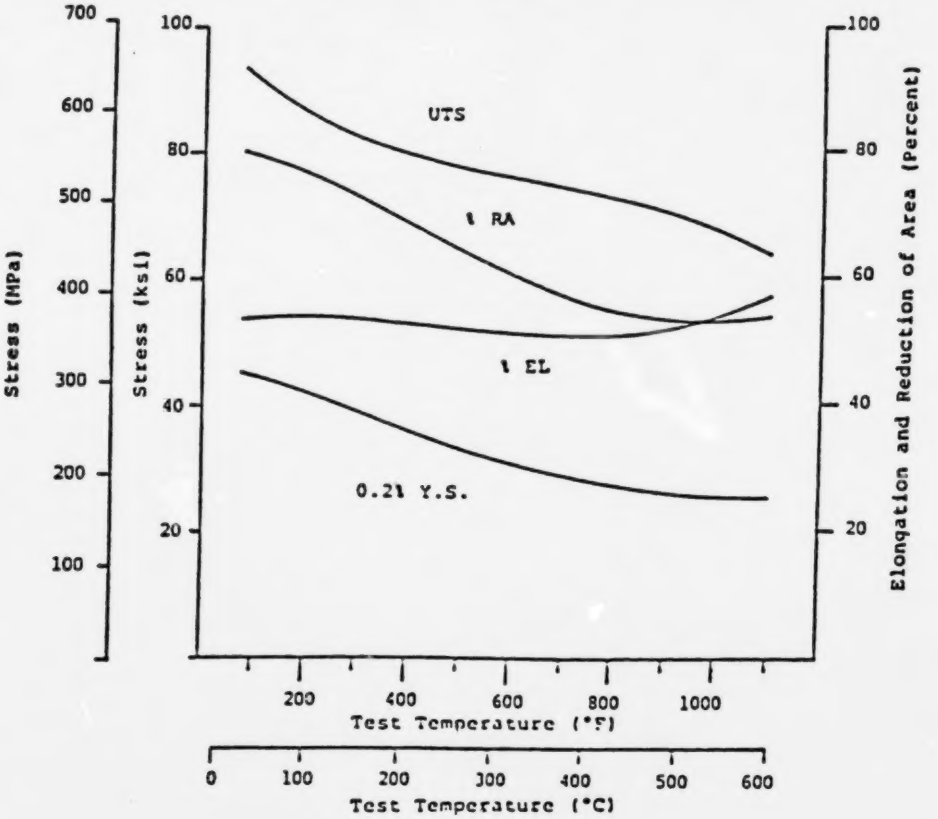


Figure 1. Elevated Temperature Tensile Properties of Carpenter 20Mo-6 Stainless (1/4" Plate-Annealed)

FABRICATION

Carpenter 20Mo-6 stainless is annealed at 1121/1204°C (2050/2200°F) for 1/2 hour per inch of thickness and water quenched. The alloy can be hardened only by cold working. There may be occasions when stress relieving is desired for complex welded structures involving 20Mo-6 stainless. Intergranular corrosion tests on welded samples suggest that stress relieving 20Mo-6 stainless is safe at temperatures below 510°C (950°F) for treating times less than 15 hours.

Alloy 625 (AWS-ERNiCrMo-3 or ENiCrMo-3) filler metal should always be used when articles of 20Mo-6 stainless are fabricated by welding. This practice is necessary so that the welds in the as-welded material have corrosion resistance equal to or better than that of the base metal. Autogenous welding is not a recommended method of fabrication. If autogenous welding is absolutely necessary, the material must be properly heat treated after welding.

Preheat and postheat are unnecessary when welding 20Mo-6 stainless. Annealing is not required after welding with Alloy 625 filler but, if performed, a water quench is suggested.

Carpenter 20Mo-6 stainless is covered by ASTM Specifications B463, B464, B468 and B474. The alloy has been approved for the ASME Boiler and Pressure Vessel Code, Code Case 1931.

SUMMARY

Carpenter 20Mo-6 stainless is a cost-effective alloy with a high level of resistance to pitting and crevice corrosion as well as other forms of corrosion. The alloy is a candidate when the pitting and crevice corrosion resistance of Hastelloy alloy G are required. Carpenter 20Mo-6 stainless possesses a combination of properties which make it attractive for a number of fume scrubber and chemical process industry applications.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

REFERENCES

1 ASTM Designation G48, "Standard Test Methods for Pitting and Crevice Corrosion Resistance of Stainless Steels and Related Alloys by the Use of Ferric Chloride Solution," Approved in 1976 and reapproved in 1980, 1983 Annual Book of ASTM Standards, Vol. 03.02, p. 280.

2 The International Nickel Co. Corrosion Engineering Bulletin No. 7, New York, p. 5 (1980).

3 Manning, P. E. CORROSION/82, NACE Meeting, Houston, Texas, Paper No. 176, March, 1982.

4 Pitler, R. K. Alloys for the Eighties, Climax Molybdenum Co., Greenwich, p. 317 (no date).

5 Dille, P. E., et. al., Seminar Entitled "Solving Corrosion Problems in Air Pollution Control Equipment," sponsored by NACE, IGCI, and APCA, Denver, Colo., Preprint No. 15, August, 1981.

6 Anderson, D. B. CORROSION/83, NACE Meeting, Anaheim, Calif., Paper No. 195, April, 1983.

7 Paul, G. T., Ross, R. W., Jr., CORROSION/83, NACE Meeting, Anaheim, Calif., Paper No. 194, April 1983.

THE CHEMISTRY OF SODIUM DRY SORBENT INJECTION

by: A.W. Mueller
A.E. Winston

Church & Dwight Co., Inc.
20 Kingsbridge Road
Piscataway, NJ 08854

ABSTRACT

Numerous studies of the dry sorbent injection method of flue gas desulfurization have been conducted and published over the past several years.

The purpose of this paper is to examine the mechanisms involved in the gas-solid reaction of sodium bicarbonate with SO_2 and thereby explain why bicarb stands out as the preferred dry sorbent.

A possible explanation of the sodium bicarbonate reaction mechanism is that the heat provided by the flue gas stream causes the thermal decomposition of sodium bicarbonate into sodium carbonate in a manner to effect optimum reaction conditions with the sulfur dioxide portion of the flue gas stream. The thermal decomposition of sodium bicarbonate over a specified temperature range is best described by a shrinking core model with chemical reaction as the rate controlling step.

As the sodium bicarbonate core decomposes, new surfaces continually become available for reaction with sulfur dioxide. The decomposition of sodium bicarbonate may also provide a moisture layer which serves as a reaction medium and thereby increases the neutralization rate. The resultant layer of sodium sulfate remains permeable so as to have no effect on reaction rate.

Data will be examined which lends support to this theory and illustrates the relationship between temperature, particle size and surface area relative to dry sorbent performance.

INTRODUCTION

BACKGROUND

Increasingly stringent regulations on emissions of SO_2 and particulate from public utility and industrial coal fired boilers make the process for simultaneous removal of these pollutants using dry sorbent injection followed by collection in a baghouse or electrostatic precipitator, attractive. Alternate system as lime/limestone wet scrubbers, dual alkali and spray drying require much greater capital and operating costs and are complex to operate.

Pilot studies on the application of dry sorbent injection utilizing fabric filters to capture particulate date back to the late sixties with studies by Wheelabrator-Frye at the Edwardsport Station of Public Service of Indiana, the Air Preheater Company at Mercer Station of Public Service Electric and Gas Company of New Jersey and at the Los Alamitos Station of Sourthern California Edison. Numerous other pilot and bench scale tests have been conducted by private industry and contracted research companies as Bechtel, KVB Corporation and various universities studies under the sponsorship of DOE, EPA and EPRI during the period from the mid-seventies to the present time.

The largest scale and most convincing demonstration to date was conducted at the 22 MW Cameo Station of the Public Service Co. of Colorado. Several tests were conducted in the 1980-82 period utilizing a number of sodium based dry sorbent materials, including conventionally mined nahcolite and trona, sesquicarbonate and soda ash. Subsequent tests have been conducted at EPRI's Arapahoe Test Facility in Denver on a 10 MW scale, expanding upon previous work testing these and other dry sorbents as commercial sodium bicarbonate, solution mined nahcolite, trona, sesquicarbonate, Wegscheiderite and light density soda ash.

The concept in all these tests is basically the same, a dry sorbent reagent is injected into the flue gas stream downstream of the air preheater, ahead of the baghouse. The reaction between the dry sorbent and the SO_2 in the flue gas stream is carried out for the most part on the surface of the bags in the baghouse. Only a limited amount of data is available utilizing this concept with an ESP in place of a baghouse.

A simplified flow diagram of this concept is illustrated in Figure 1 below:

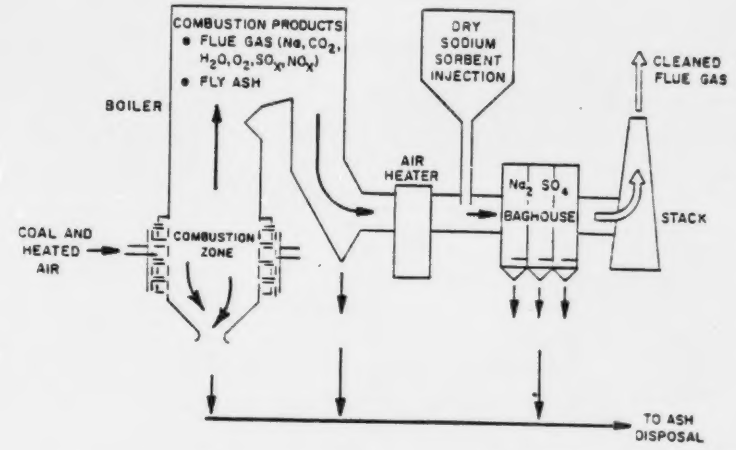

COMBUSTION PRODUCTS
• FLUE GAS (N_2, CO_2, H_2O, O_2, SO_x, NO_x)
• FLY ASH
BOILER
DRY SODIUM SORBENT INJECTION
CLEANED FLUE GAS
AIR HEATER
Na_2SO_4
BAGHOUSE
STACK
COAL AND HEATED AIR
COMBUSTION ZONE
TO ASH DISPOSAL

Figure 1. Schematic of a coal-fired power plant with dry-sodium injection for SO_2 control.

In each comparative test it has been recognized that sodium bicarbonate in the manufactured or mineral form is the most effective sorbent tested followed by trona or sodium sesquicarbonate and soda ash. This is illustrated in Figure 2 below.

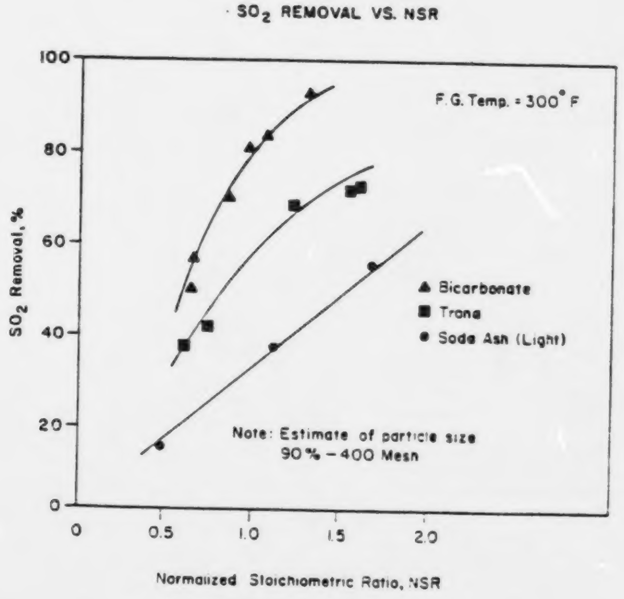


Figure 2. SO_2 removal vs NSR for three reagents; sodium bicarbonate, trona and light density soda ash

The non-sodium compounds as lime and its hydrates, limestone, and dolomite are less effective and accordingly have received less attention in recent studies.

Of the sodium compound the relative effectiveness can be explained by simplistically by a mechanism whereby sodium bicarbonate thermally decomposes in the flue gas stream in the temperature range of 250-350°F while simultaneously undergoing a reaction with SO_2 to form sodium sulfate. The sulfate is preferentially formed due to the presence of excess oxygen in the flue gas stream.

ADVANTAGES OVER ALTERNATE SYSTEMS

Dry sodium injection inherently offers a number of advantages over alternate FGD systems, some of which have been mentioned above. These include:

- effective removal of $\geq$ 90% SO_2 at near stoichometric levels
- lower capital and operating costs
- simplicity of operation
- easily retrofitted to boilers equipped with baghouse or possibly ESP's
- ease of transporting of dry solids waste
- lower volume of solids waste
- lower power costs

The major disadvantages voiced in the past were sodium sorbent reagent availability and the problem of soluble sodium waste. The first of these is no longer a problem since a number of reagent suppliers offer various forms of sodium sorbents from refined sodium bicarbonate and solution mined nahcolite to trona and light density soda ash. The volumes available now and mineral reserves will easily meet current needs and those well into the future.

The problem of soluble sodium waste has been the subject of past studies and of current studies and a number of solutions are possible depending on the region of the U.S.. In more arid climates a simple landfill operation may be possible. In most areas of the country however, there is a net positive rainfall and the leaching of soluble salts must be considered. Various solutions have been investigated including:

- clay isolation cells
- pelletizing/sintering
- encapsulation
- vitrification
- chemical fixation
- sodium recovery and use in other industries
- sodium regeneration
- ocean dumping
- deep well injection

Some of these will be acceptable in certain areas and others will not depending on environmental consideration and economics.

The subject of this paper is to examine the mechanics of the sodium dry sorbent reaction and relate the factors involved to the observation from the various studies that sodium bicarbonate reacts more completely than the other sodium reagent forms. Specifically, the three areas which will be examined are:

- the thermal decomposition of sodium bicarbonate and its reaction with SO_2
- the effect of particle size of sodium bicarbonate on the reaction with SO_2
- the relationship of surface area to particle size

THERMAL DECOMPOSITION OF SODIUM BICARBONATE AND REACTION WITH SO_2

As mentioned in the introduction numerous studies have shown that sodium bicarbonate reacts more completely with SO_2 in the flue gas stream than the other candidate materials as trona, sodium sesquicarbonate or sodium carbonate. An explanation for this is that the heated flue gas thermally decomposes the $NaHCO_3$ into Na_2CO_3 which simultaneously reacts with SO_2.

The decomposition reaction for sodium bicarbonate is given by

$$2NaHCO_3(s) \xrightarrow{\Delta} Na_2CO_3(s) + CO_2(g) + H_2O(g)$$

The temperature at which decomposition occurs will depend on the partial pressure of CO_2 and water vapor in the flue gas.

The thermal decomposition temperature is an important factor in pore development of the product sodium carbonate. The optimum decomposition temperature range determined by several studies (Keener, Davis 1984) appears to be 225-350°F depending on the particle size used. Above 600°F a loss of specific surface area has been observed which was attributed to sintering.

Thermal decomposition data fractional conversion versus time for three particle size ranges and six temperatures in the range 225-350°C were generated in lab apparatus (Keener, Davis 1984). This data is given in Figures 3 and 4 on the following page. The temperature range was selected to cover that which might be experienced in fabric filter operation on flue gas streams.

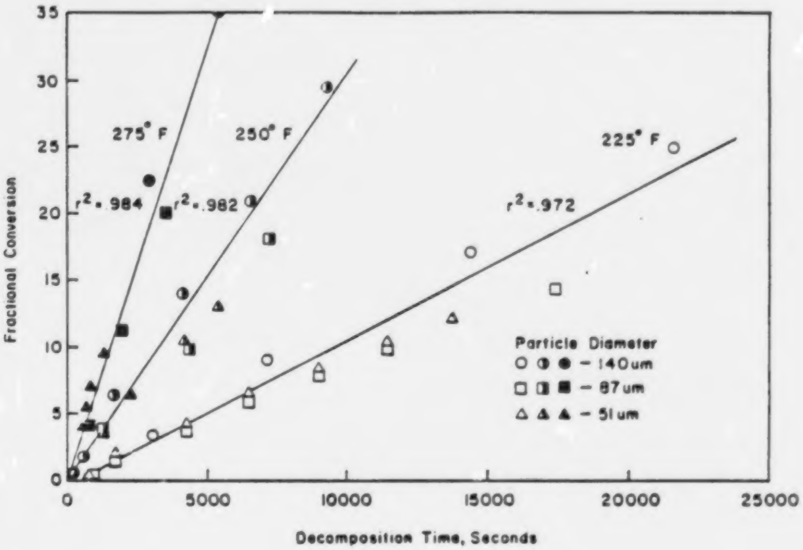


Figure 3. Decomposition of sodium bicarbonate in the temperature range of 225-275°F

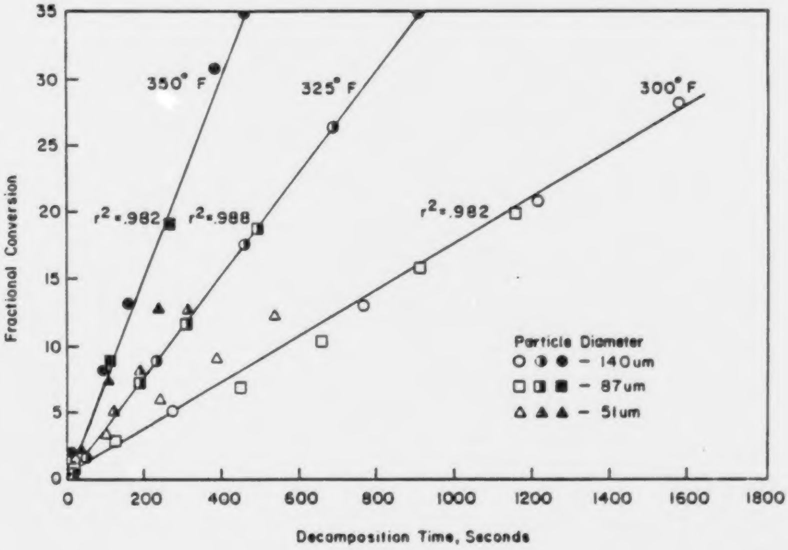


Figure 4. Decomposition of sodium bicarbonate in the temperature range of 300-350°F

This data was fit to a shrinking core model assuming that thermal decomposition reactions occur at interfaces within the solid particle. The rate of movement of this interface from the outer edge of the decomposing solid toward its center may be controlled by; heat transfer from the gas stream to the solid, the thermal properties of the solid, the decomposition reaction rate or the rate of diffusion of the decomposition gaseous products away from the interface. In other words, the decomposition rate may be controlled by either heat transfer, chemical reaction or diffusion.

From previous studies on calcium carbonate and from the decomposition data on sodium bicarbonate it was concluded that chemical reaction is the rate controlling step and not heat transfer or diffusion.

The shrinking core model is developed using the assumption that the reaction rate is proportional to the reactant/solid product interfacial surface area. If the fractional conversion is plotted versus the time as in Figures 3 and 4 the slope k or reaction rate coefficient can be determined for each temperature. A plot of the natural log of the rate coefficients k versus the reciprocal temperatures yields a straight line giving an equation of the Arrhenius form:

$$k = Ae^{-E/RT}$$

The values for activation energy, E, and the constant A are 20.5 kcal/mole and 4.91×10^4 g/cm^{2}s, respectively

These results strongly support the assumption that the decomposition reaction is kinetically controlled over the temperature range investigated (225-350°F), (Keener, Davis 1984).

Also from this one can derive the relationships for conversion time, t, or complete conversion (decomposition) τ, of a $NaHCO_3$ particle of radius R_s.

$$\tau = \frac{\rho_A}{a\,k}\frac{Mc}{M_A}R_s$$

where

Mc = molecular weight of $NaHCO_3$
M_A = molecular weight of CO_2
ρ_A = density of $NaHCO_3$ g/cm3
R_s = initial radius of NaHCO3, cm
a = stoichiometric coefficient, 2 moles $NaHCO_3$/mole CO_2
k = reaction rate coefficient g/cm2s

This equation can be used for estimating decomposition times over the temperature range of the model (225-350°F).

The conclusions derived from this work are:

- The thermal decomposition of $NaHCO_3$ is chemical kinetically controlled in the temperature range of 225-350°F.
- A shrinking core model provides a good fit to the data. The rate coefficient, k, is of the Arrhenius form with an activation energy of 20.5 kcal/mole.
- The time for total decomposition of a sodium bicarbonate particle can be estimated by use of the formula:
  $\tau = \rho \ \ M_c R_s/(a\ k\ M_A)$
- Water vapor and CO_2 partial pressures normally found in flue gases should not affect sodium bicarbonate decomposition rates above 300°F.

The above discussion of the rate of decomposition of sodium bicarbonate is of interest because of its influence on the rate of reaction between $NaHCO_3$ and SO_2 in flue gas.

In the flue gas stream the reaction of sodium bicarbonate with SO_2 can be described by the following equations.

$$2NaHCO_3(s) \longrightarrow Na_2CO_3(s) + H_2O(g) + CO_2(g) \qquad (1)$$

$$2NaHCO_3(s) + SO_2(g) \longrightarrow Na_2SO_3(s) + 2CO_2(g) + H_2O(g) \qquad (2)$$

$$Na_2CO_3(s) + SO_2(g) + 1/2\ O_2(g) \longrightarrow Na_2SO_4(s) + CO_2(g) \qquad (3)$$

$$Na_2CO_3(s) + SO_2(g) \longrightarrow Na_2SO_3(s) + CO_2(g) \qquad (4)$$

$$Na_2SO_3(s) + 1/2\ O_2(g) \longrightarrow Na_2SO_4(s) \qquad (5)$$

The first reaction occurs much faster than reaction (2). This reaction produces a more porous particle with a greater reacting surface. Thus, it is probably reaction (3) or (4) which is responsible for SO_2 removal.

Once again a shrinking core model can be used to describe the reaction. There are three resistances that could determine the rate of reaction and if any one is significantly slower than the others it controls the rate of reaction. These steps are: (1) gas phase resistance, diffusion of SO_2 molecules to the particle surface from the gas stream (2) product layer diffusion resistance, diffusion of the SO_2 molecules through the Na_2SO_4 layer around the Na_2CO_3 and (3) reaction rate at the product/ reactant interface.

Since increased efficiency of the SO_2 reaction with $NaHCO_3$ is observed with increasing temperature over the range 225-350°F (see Figure 5) it can be assumed that chemical reaction is the rate controlling step.

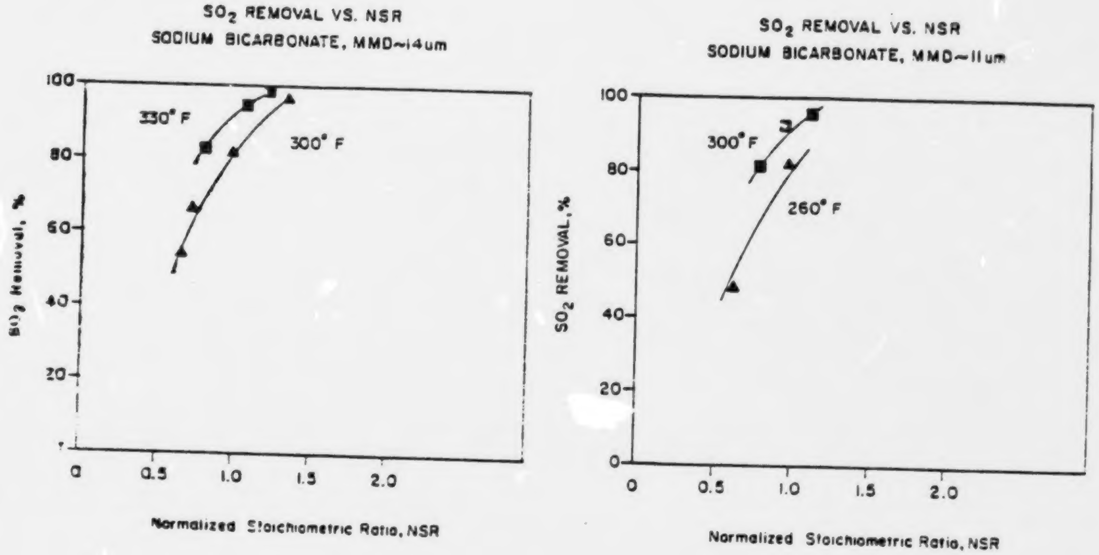


Figure 5. SO_2 removal versus NSR for two sodium bicarbonate reagents illustrating the dependence of SO_2 removal to flue gas temperature

Above 350°F the thermal decomposition of $NaHCO_3$ begins to occur too fast in comparison to the reaction with SO_2 and the SO_2 removal efficiency drops off. This is illustrated in Figure 6. It is possible that the presence of water provided by $NaHCO_3$ decomposition, serves a role in maximizing the SO_2 neutralization rate. Thus at moderate temperatures a very thin layer of moisture may be maintained on the reacting surfaces. Under these conditions SO_2 neutralization would be a rapid solution mediated reaction. With higher temperatures, however, water loss may be too rapid giving rise to slower gas/solid state reactions. This theory is somewhat supported by noting that at 450°F and above the bicarbonate efficiencies once again begin to increase with temperature and become similar to those obtained with Na_2CO_3. Above 650°F both $NaHCO_3$ and Na_2CO_3 show a tendency to sinter, further decreasing the efficiency of the conversion to sodium sulfate.

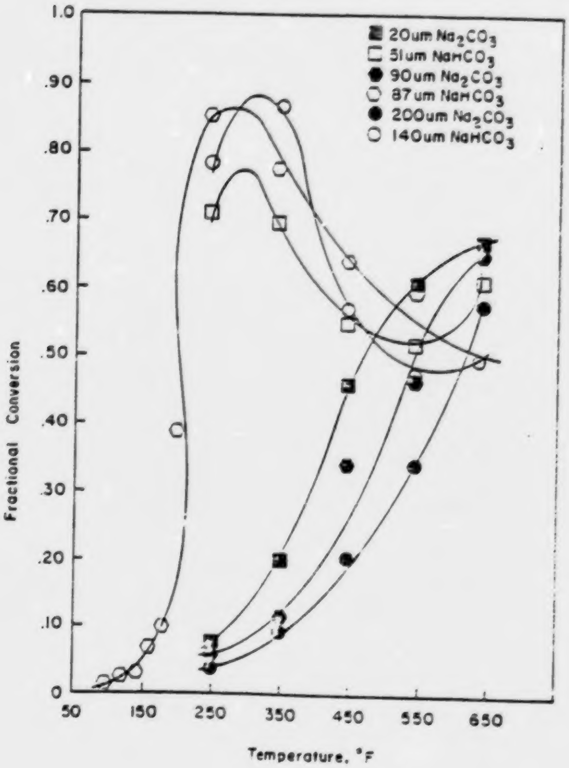


Figure 6. Comparison of conversion Values for Na_2CO_3 and $NaHCO_3$

Thus, two simultaneous reactions are occurring in the flue gas stream; the thermal decomposition of sodium bicarbonate reaction (1) and the reaction of the resulting sodium carbonate with SO_2, either reaction (3) or (4) or both.

It is hypothesized that these two reactions occur at approximately the same rate for a temperature range of 225-350°F, thus accounting for the high rate of conversion of sodium bicarbonate and SO_2 removal in the flue gas stream. The overall reaction mechanism is illustrated in Figure 7.

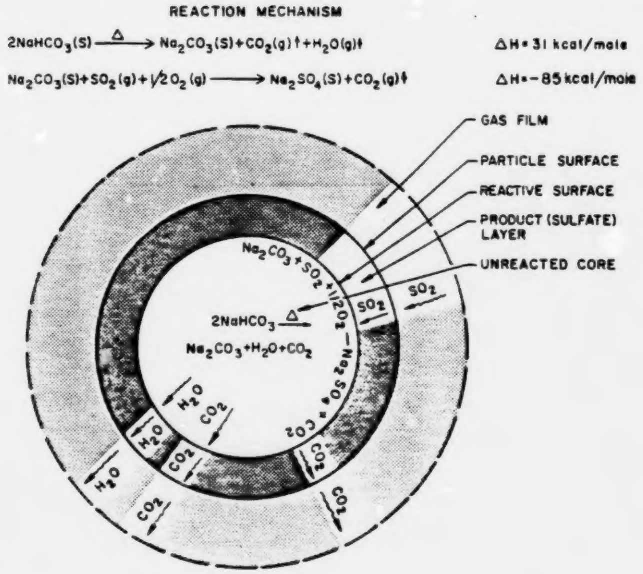

REACTION MECHANISM
2NaHCO3(S) Δ→ Na2CO3(S)+CO2(g)↑+H2O(g)↑
ΔH=31 kcal/mole
Na2CO3(S)+SO2(g)+1/2O2(g) → Na2SO4(S)+CO2(g)↑
ΔH=-85 kcal/mole
GAS FILM
PARTICLE SURFACE
REACTIVE SURFACE
PRODUCT (SULFATE) LAYER
UNREACTED CORE
Na2CO3+SO2+1/2O2—Na2SO4+CO2
2NaHCO3 Δ→ Na2CO3+H2O+CO2
SO2
SO2
H2O
CO2
H2O
CO2
CO2
CO2
H2O
CO2
CO2

Figure 7. Representation of reacting sodium bicarbonate particle

Variation in conversion efficiencies are likely caused by temperature and particle size considerations or physical factors as the distribution on the fabric filter and operating mode of the baghouse. The next section will examine data on the effect of particle size on conversion efficiencies.

EFFECT OF $NAHCO_3$ PARTICLE SIZE IN REACTION WITH SO_2

The most recent and comprehensive work on effect of $NaHCO_3$ particle size on reaction with SO_2 in the flue gas stream was conducted by EPRI and KVB Inc. at the Arapahoe test station in Denver, Colorado in 1984.

Results of this test work are illustrated in Figure 8 below:

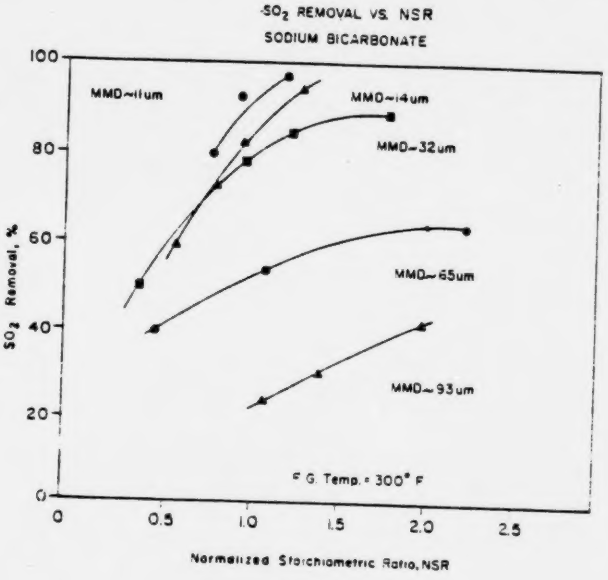


Figure 8. SO_2 removal vs NSR for sodium bicarbonate illustrating the dependence of SO_2 removal on particle size MMD (Mass Mean Diameter)

SO_2 removal is shown to significantly improve with decreasing sodium bicarbonate particle size. Removal of greater than 90% of the SO_2 is possible with the finer particle size (MMD of 11 micron).

This behavior would be expected from the mechanism described above. Since the chemical reaction is the rate controlling step, the smaller particle will react faster and more completely because of their greater specific surface area.

However, this is further enhanced by the physical distribution of these fine particles on the baghouse filter media. Typically 70-80% of the SO_2 removal takes place on the filter media in the baghouse and 20-30% of the remaining SO_2 removal is in the ductwork. Typical utility baghouses are in-out designs where the gas enters a plenum above the dust hopper and passes through a tube sheet upon which the bags are attached. The particulate material is deposited on the inside of the bags and is removed by periodically reversing the gas flow thus causing the cake formation on the inside of the bags to fall off into the dust collector hopper.

In the case where particles are of marginal size to remain fluidized at the entry to the baghouse a portion will settle into the dust hopper or deposit non-uniformity on the bags. This will reduce the reagent utilization.

This emphasizes the importance of maintaining transport velocity of the particles in the flue gas stream from the injection point to where they impinge on the filter cloth media in the baghouse. This is of particular concern at lower boiler steam loads where flue gas flow rates are substantially lower than typical operation.

RELATIONSHIP OF SURFACE AREA TO PARTICLE SIZE

In the previous section it was shown that SO_2 removal efficiencies improve dramatically with smaller sodium bicarbonate size. For material of mass mean diameter (MMD) of 10 microns removal of greater than 90% is possible. It is an obvious assumption that as the particles become smaller the specific surface area increases, and this is the case as can be seen in Figure 9. The change is small for average particle sizes of 80 to 40 micron. Below 40 micron the slope of the curve changes and the specific surface area increases dramatically. No data is available below 10 micron.

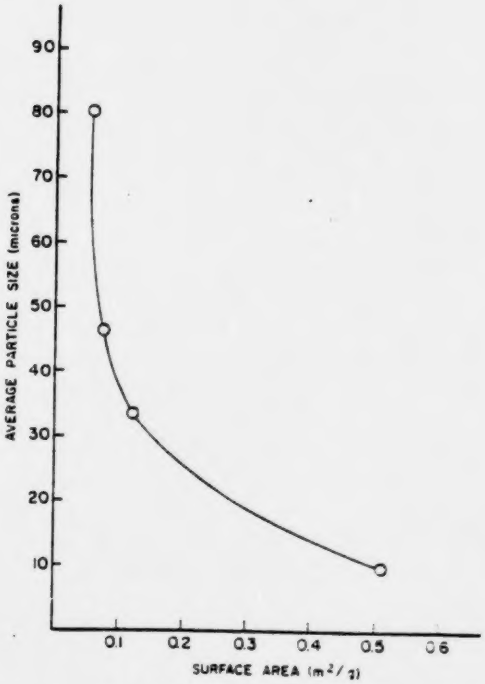


Figure 9. Average particle size (microns) versus surface area (m2/g)

At 10 micron average particle size sodium bicarbonate is essentially completely utilized. At a stoichiometric ratio of 1 or slightly greater, better than 90% of the SO_2 can be removed. It is unlikely that greater specific surface ares would in any way improve utilization of sodium bicarbonate reagent or benefit SO_2 removal efficiency.

As the sodium bicarbonate decomposes CO_2 and water vapor are diffused through the unreacted sodium bicarbonate particle core and sulfate product layer exposing new surface in the process. SO_2 from the flue gas stream diffuses through the particle and reacts on the product/reactant interface surface.

The specific surface area during thermal decomposition of sodium bicarbonate increases by a factor of as high as 30 from initial to complete conversion as shown in Figure 10.

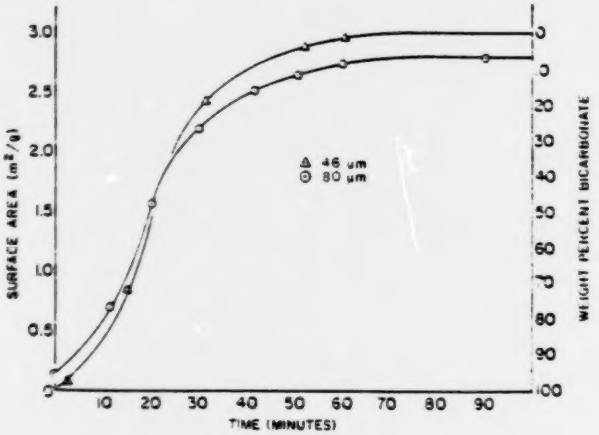


Figure 10. Thermal decomposition of sodium bicarbonate versus time surface area for sodium bicarbonate of 46 and 80 microns average particle size

SODIUM BICARBONATE SURFACE AREA FROM PARTICLE SIZE MEASUREMENT

Surface area of sodium bicarbonate can be calculated from the particle size distribution and by assuming a particle shape. As illustrated in Table I surface area is dependent on particle shape. If the L/D ratio changes from 1:1 to 10:1 the corresponding surface area will be reduced by 30%.

TABLE I

DEPENDENCE OF SURFACE AREA ON PARTICLE SHAPE (PARTICLE SIZE 115 MICRONS (-120, +140 MESH) CUT, CRYSTAL DENSITY 2.2g/cm^3

	PARTICLE SHAPE				
	SPHERE	CUBE	L/D=2/1	L/D=5/1	L/D=10/1
VOLUME=V_1 (cm^3)	$7.963 \cdot 10^{-7}$ ($\frac{\pi}{6} D^3$)	$1.521 \cdot 10^{-6}$ (d^3)	$3.042 \cdot 10^{-6}$ ($2d^3$)	$7.604 \cdot 10^{-6}$ ($5d^3$)	$1.521 \cdot 10^{-5}$ ($10d^3$)
WEIGHT=W_1 (g.)	$1.752 \cdot 10^{-6}$ ($2.2 \cdot V_1$)	$3.346 \cdot 10^{-6}$ ($2.2 \cdot V_1$)	$6.692 \cdot 10^{-6}$ ($2.2 \cdot V_1$)	$1.673 \cdot 10^{-5}$ ($2.2 \cdot V_1$)	$3.346 \cdot 10^{-5}$ ($2.2 \cdot V_1$)
SURFACE AREA= S_1 (cm^2)	$4.155 \cdot 10^{-4}$ (πD^2)	$7.935 \cdot 10^{-4}$ ($6d^2$)	$1.322 \cdot 10^{-3}$ ($10d^2$)	$2.909 \cdot 10^{-3}$ ($22d^2$)	$5.554 \cdot 10^{-3}$ ($42d^2$)
NUMBER OF PARTICLES=N_T ($\frac{\#}{g.}$)	570802 ($\frac{1}{W_1}$)	298864 ($\frac{1}{W_1}$)	149436 ($\frac{1}{W_1}$)	59774 ($\frac{1}{W_1}$)	29887 ($\frac{1}{W_1}$)
TOTAL SURFACE AREA=S_T ($\frac{m^2}{g.}$)	0.0237 ($N_T \cdot S_1$)	0.0237 ($N_T \cdot S_1$)	0.0198 ($N_T \cdot S_1$)	0.0174 ($N_T \cdot S_1$)	0.0166 ($N_T \cdot S_1$)

Reproduced from best available copy.

Table 2 illustrates how surface area varies with particle size from 149 to 1 micron assuming a cubic particle shape. For an average cut diameter (ACD) of 163 to ACD of 2 the surface area increases by a factor of 80.

TABLE 2

DEPENDENCE OF SURFACE AREA ON PARTICLE SIZE

(PARTICLE SHAPE = CUBE)

(CRYSTAL DENSITY = 2.2g./cm^3)

U.S.S. MESH	SIZE, MICR.	AVERAGE CUT DIAMETER MICR.	ONE PARTICLE: VOLUME V_1(cm^3) (d^3)	ONE PARTICLE: WEIGHT W_1(g.) ($2.2V_1$)	ONE PARTICLE: SURFACE AREA S_1(cm^2) ($6d^2$)	PARTICLES PER GRAM: NUMBER ($N_T$per gram) ($\frac{1}{w_1}$)	PARTICLES PER GRAM: SURFACE AREA S_T(m^2/g.) ($N_T.S_1$)
100	149	163	$4.330\cdot10^{-6}$	$9.530\cdot10^{-6}$	$1.594\cdot10^{-3}$	104958	0.0167
120	125	137	$2.571\cdot10^{-6}$	$5.657\cdot10^{-6}$	$1.126\cdot10^{-3}$	176773	0.0199
140	105	115	$1.521\cdot10^{-6}$	$3.346\cdot10^{-6}$	$7.935\cdot10^{-4}$	298871	0.0237
170	88	96	$8.847\cdot10^{-7}$	$1.946\cdot10^{-6}$	$5.530\cdot10^{-4}$	513764	0.0284
200	74	81	$5.314\cdot10^{-7}$	$1.169\cdot10^{-6}$	$3.937\cdot10^{-4}$	855307	0.0337
230	63	68	$3.144\cdot10^{-7}$	$6.917\cdot10^{-7}$	$2.774\cdot10^{-4}$	$1.44\cdot10^{6}$	0.0401
270	53	58	$1.957\cdot10^{-7}$	$4.292\cdot10^{-7}$	$2.018\cdot10^{-4}$	$2.33\cdot10^{6}$	0.0470
325	44	48	$1.106\cdot10^{-7}$	$2.433\cdot10^{-7}$	$1.382\cdot10^{-4}$	$4.11\cdot10^{6}$	0.0568
400	37	40	$6.400\cdot10^{-8}$	$1.408\cdot10^{-7}$	$9.600\cdot10^{-5}$	$7.10\cdot10^{6}$	0.0682
425	33	35	$4.287\cdot10^{-8}$	$9.432\cdot10^{-8}$	$7.350\cdot10^{-5}$	$1.06\cdot10^{7}$	0.0779
500	25	29	$2.439\cdot10^{-8}$	$5.365\cdot10^{-8}$	$5.046\cdot10^{-5}$	$1.86\cdot10^{7}$	0.0940
625	20	22	$1.064\cdot10^{-8}$	$2.342\cdot10^{-8}$	$2.904\cdot10^{-5}$	$4.27\cdot10^{7}$	0.1240
800	15	17	$4.913\cdot10^{-9}$	$1.081\cdot10^{-8}$	$1.734\cdot10^{-5}$	$9.25\cdot10^{7}$	0.1604
1250	10	12	$1.727\cdot10^{-9}$	$3.801\cdot10^{-9}$	$8.640\cdot10^{-6}$	$2.63\cdot10^{8}$	0.2272
2500	5	7	$3.430\cdot10^{-10}$	$7.546\cdot10^{-10}$	$2.940\cdot10^{-6}$	$1.32\cdot10^{9}$	0.3896
6850	2	3	$2.700\cdot10^{-11}$	$5.940\cdot10^{-11}$	$5.400\cdot10^{-7}$	$1.68\cdot10^{10}$	0.9091
12500	1	2	$8.000\cdot10^{-12}$	$1.760\cdot10^{-11}$	$2.400\cdot10^{-7}$	$5.68\cdot10^{10}$	1.3636

From calculation of the numbers in Table 2 it was observed that the surface area in m^2/g for a given average cut diameter could be obtained by using

$$S_T\ (m^2/g) = \frac{2.727}{ACD\ (\text{microns})}$$

Accordingly, only a detailed particle size analysis is required for a determination of total surface area of a given sample. For example in Table 3 a commercial grade sodium bicarbonate (Church & Dwight 3DF), surface area is calculated from a particle size distribution obtained on an Elzone analyzer and compared to actual measured BET surface area.

TABLE 3

THEORETICAL CALCULATION OF SURFACE AREA FOR 3DF (ELZONE PARTICLE SIZE ANALYSIS). PARTICLE SHAPE ASSUMED TO BE CUBE.

AVERAGE CUT DIAMETER, MICRONS	DIFFERENTIAL WT. %	SURFACE AREA CONTRIBUTION, m^2 $\left(\frac{2.727}{ACD} \cdot \text{Diff. Wt.\%}\right)$
5	.26	0.0014
8	2.19	0.0075
10	2.93	0.0080
12	3.58	0.0082
14	3.87	0.0076
16	4.14	0.0071
18	3.83	0.0058
20	4.20	0.0057
25	10.62	0.0016
30	11.01	0.0100
35	11.04	0.0086
40	12.16	0.0083
45	10.48	0.0064
50	7.86	0.0043
55	5.69	0.0028
60	2.71	0.0012
70	1.13	0.0009
80	1.24	0.0004

TOTAL CALCULATED SURFACE AREA = 0.106 m^2/g.

MEASURED BET SURFACE AREA = 0.202 m^2/g.

This procedure was repeated for two other commercial grade sodium bicarbonates and compared to measured numbers Table 4. Since in each case the measured surface area is 2-3.5 times greater than the calculated this figure was designated as a roughness factor to be applied to the calculated numbers.

TABLE 4

ROUGHNESS FACTOR CALCULATION: $R_f = \frac{\text{Measured Surface Area}}{\text{Calculated Surface Area}}$

MATERIAL	SURFACE AREA, m^2/g CALCULATED	SURFACE AREA, m^2/g BET MEASURED	ROUGHNESS FACTOR, R_f
Sodium Bicarbonate C&D Grade 5	.020	.069 (.086)	∿ 3.5
Sodium Bicarbonate C&D Grade 3DF	0.106	0.202	∿ 2.5
Sodium Bicarbonate	.057	(Est.) 0.140	(Est.) 2.5

From this work one can conclude that surface area is a function of particle shape as well as distribution and that material described by average particle size or mass mean diameter alone can vary considerably in surface area.

SUMMARY

Impending legislation on SO_2 control and new requirements for particulate control based on fine (10 micron and below particles) will make the process for simultaneous removal utilizing dry sorbent injection followed by a baghouse or electrostatic precipitator an attractive alternative.

Numerous pilot studies have substantiated the viability of this technology up to a scale of 22 MW. The concept in all these tests is basically the same, a dry sorbent reagent is injected into the flue gas stream downstream of the air preheater ahead of a baghouse. The reaction between the dry sorbent reagent and the SO_2 in the flue gas stream is carried out on the surface of the filter media in the baghouse.

In all tests it has been recognized that sodium bicarbonate is the most effective dry sorbent tested. This can be explained by a mechanism whereby sodium bicarbonate thermally decomposes in the flue gas stream while simultaneously undergoing a reaction with SO_2 to form sodium sulfate. Decomposition of the sodium bicarbonate serves to increase the surface area and perhaps to provide some moisture as a reaction medium.

Over the temperature range 225-350°F, it can be assumed that the chemical reaction is the rate controlling step. Above 350°F the thermal decomposition of $NaHCO_3$ begins to occur too fast in comparison to the SO_2 neutralization reaction and the SO_2 removal efficiency drops off.

From the data (Keener, Davis 1984) a shrinking core model was developed for decomposition. Likewise the reaction of sodium carbonate with SO_2 to product sodium sulfite or sulfate can be depicted by a shrinking core model.

Recent work conducted by EPRI in the latter part of 1984 shows the dependence of SO_2 removal on particle size of sodium bicarbonate. SO_2 removal was shown to significantly improve with decreasing sodium bicarbonate particle size. Removal of greater than 90% of the SO_2 is possible with material $\sim$ 10 micron MMD.

The improved performance observed as the average particle size of the dry sorbent reagent decreases is related to surface area. For average particle size of 80 to 40 micron the change in specific surface area is relatively small. Below 40 micron down to 10 micron the specific surface area increases at a much greater rate. At a 10 micron average particle size sodium bicarbonate is essentially completely utilized. At a stoichiometric ratio of 1 or slightly greater 90% of the SO_2 can be removed. It is unlikely that greater surface areas could improve the sodium bicarbonate reagent removal efficiency.

Surface area of sodium bicarbonate can be calculated from the particle size distribution and assuming a particle shape. This calculated surface area gives reasonable results but must be multipled by a factor to agree with measured BET surface area. In deriving an expression for surface area and observing corresponding measured data it can be concluded that since surface area is a function of particle size distribution and shape that material described by average particle size alone can vary considerably in surface area.

Physical consideration on the baghouse design are important. These fine particles are less likely to settle out in the dust hopper and distribute more evenly on the filter media.

The advantages of dry sodium sorbent injection are numerous; effective removal of $\geq$ 90% SO_2 at near stoichiometric levels, low capital and operating costs, easily retrofitted and lower volume of solids waste, readily available supply of reagents as well as future reserves.

The only disadvantage recognized is the problem of soluble waste. This is not a problem in an arid area and numerous solutions are possible in other areas from clay isolation cells to chemical fixation or even regeneration.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

REFERENCES

1. Keener, T.C., Frazier, G.C., and Davis, W.T. Thermal Decomposition of Sodium Bicarbonate in: Chemical Engineering Communications Vol. 33, Pg 93-105, June 1984.

2. Keener, T.C., Davis, W.T. Study of the Reaction of SO_2 with $NaHCO_3$ and Na_2CO_3 in: Journal of the Air Pollution Control Association, June 1984, Volume 34, Number 6, JPCAAC 34(6)611-708(1984).

3. Bland, V.V., Hooper, R.G., Characterization of Alternate Sodium Sorbents for Fabric Filter SO_2 Capture, Presented at the First Joint Symposium on Dry SO_2 and Simultaneous SO_2/NO_x Control Technologies.

4. Davis, W.T., Keener, T.C., Chemical Kinetic Studies on Dry Sorbents - Final Report, DOE/FC/10184-1, U.S. Department of Energy, National Technical Information service, February 1982.

TECHNO-ECONOMIC INVESTIGATION INTO THE HYBRID TECHNOLOGY OF FGD AND PCC APPLIED TO LARGE BOILERS OF FIRING CANADIAN COALS

By: R. L. Wang, Ph. D.; P. E.
Engineering Consultant
St. Catharines, Ontario, Canada L2T 2J2

ABSTRACT

Under the policy of energy self-sufficiency in Canada, the federal government has encouraged large boiler users to fully utilize indigenous coals as a primary fuel source for steam generation.

At the same time, the public awareness of adverse impact of acid rain on the northern environment will lead the present SO_2 emission control guideline to be further stringent. Utility users, engineering designers, science investigators and others are paying their attention to the study of operationally reliable and cost saving measures for removal of sulfur compounds anterior and posterior to combustion (in some cases even dealing with in-suit capture in a combustion region).

This paper exames coal properties of major Canadian reserves required for utility/industrial steam generation. Main commercialized technologies of FGD (Flue Gas Desulfurization) and PCC (Physical Coal Cleaning) are reviewed. Their cost accounts based on individual and hybrid systems are estimated. Results of cost benefit analysis on capital investments, annual operating /maintenance cost and equivalent unit revenue requirement in life cycle operation are presented.

INTRODUCTION

The world industrialized societies are responsible for massive amounts of sulfur dioxide pollutant being discharged into the atmosphere each year - about 65 million metric tons. Three quarters of that amount comes from one percent of the global surface area, namely the central and north eastern United States, southern Canada, northern Europe and Japan. Increasing worldwide demand for energy, derived mainly from burning coals, will make this situation even worse by the twenty-one century.

Sulfur dioxide, besides causing human respiratory irritation and corrosion on structural properties, has been found to be a major source of acid rain formation, when it oxides into acidic sulfate (H_2SO_4) in the presence of moisture in an activiated atmosphere. Since World War II, the average acidity of precipitation in the eastern Canada, the northeastern United States and Scandinavia has increased significantly.

Although the impacts of acid precipitation have not been fully evaluated, studies in Europe, Canada, the U.S.S.R. and the northeastern United States have shown a number of adverse effects on the damage of ecosystems. It is believed that acid rain causes the declination of the present forest and agricultural productivity.

Due to public awareness of such adverse effects, the Canadian National Guidelines for emissions from stationary sources, especially for new units, have become more stringent. As a result of encouragement of burning indigenous coals for future large boilers under the policy of energy self-sufficiency in Canada with simultaneous upgrading of emission guidelines, utility users, engineering designers, science investigators and others have already paid their attention to study operationally reliable and cost saving measures for removal of sulfur compounds anterior and posterior to combustion. In this paper, a techno-economic study is made on individual and hybrid systems of FGD and PCC for large coal-fired boilers.

REVIEW OF CANADIAN NATIONAL GUIDELINES FOR SO_2 EMISSION CONTROL

The Evironment Canada, a federal department, issued National Guidelines for New Stationary Scources, in 1981, indicating the quantities and concentration above which SO_2, NO_x and particulate matter should not be emitted from fossil fuel-fired steam-driven electricity power plants. The guidelines are applied to new units only.

For sulfur dioxide, generating units emitting between 258 ng/J and 2580 ng/J of heat input, when uncontrolled, should be controlled so that their final emission does not exceed 258 ng/J. Units emitting over the margin of 2580 ng/J should be controlled so that a minimum of 90% of the uncontrolled emission is captured before release to the atmosphere.

The permitted emissions are based on compliance of monitored 720-hour rolling emission average which excludes intervals of shut-down hours. The emission limits recommended are achievable using demonstrated and commercial control methods. Technologies discussed in this paper is, therefore, evaluated having regard to compliance with these guidelines.

COAL RESERVES IN CANADA AND THEIR PROPERTIES

Canadian coal reserves distribute mainly in two major geographic regions, namely, British Columbia and Alberta in the western region, and New Brunswick and Nova Scotia in the eastern region. Southern Saskatchewan is the area of rich deposit of lignite. The quantities of measured coal resources of immediate interest in Canada are estimated about 16.53 giga tonnes of bituminous coal and 3.54 giga tonnes of lignite as mentioned in the Report ER 79-9, Department of Energy, Mines and Resources.

Generally, western coals are categorized as low sulfur and low volatile fuels and eastern coals as high sulfur and high volatile ones. For the sake of simplicity in economic comparison on an equivalent base, the type of bituminous coals is selected in the study as a reference fuel for techno-economic evaluation. Analytical data published in References (1) and (2) are used to define the range of sulfur and thermal properties of the eastern and western bituminous coals.

The shadowed areas in Fig 1 represent the two regions of interest in the study. A group of lines in the upper portion of the figure represents the relationship between coal sulfur content and potential SO_2 generation from coals provided that no

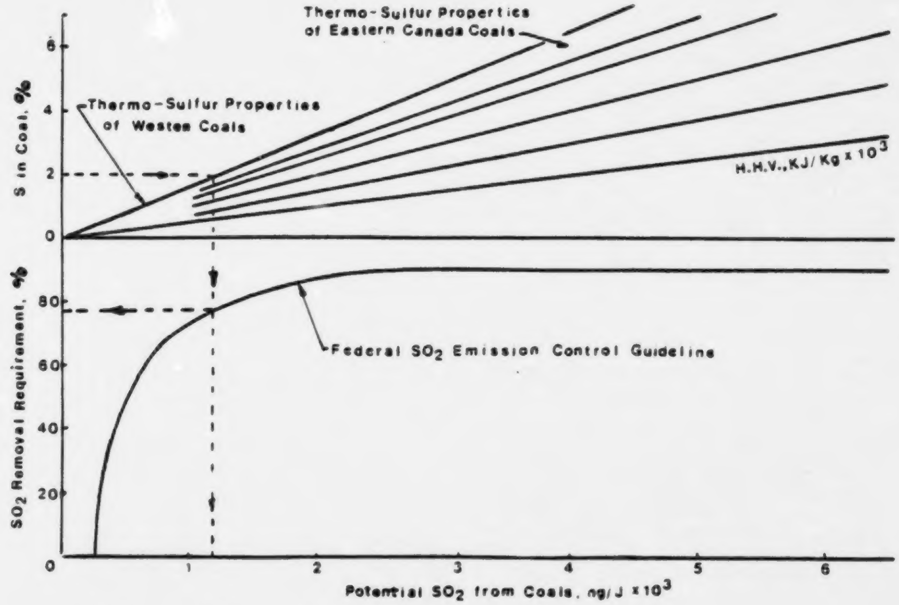


Fig 1 SO_2 Removal Requirement Coorelated with Coal H.H.V.

sulfur retains in refuse after the completion of coal combustion. Coal higher heating value in KJ/Kg is indicated for each line. Analytical data from Tibbetts and Montgomery's papers (1), (2) reveal that eastern coals have wider range of sulfur contents as compared with those of western coals.

The lower portion of the figure refers to the Canadian federal guideline for SO_2 emission control. Above the emission level of 2580 ng/J, the guideline curve becomes horizontal at 90% of SO_2 removal requirement. From a combination of the two portions, a required SO_2 removal can be found for a given S content in coal and a coal H.H.V. For example, for a 2% S and 32000 KJ/Kg H.H.V. coal, the emission level and percentage of removal requirement are found to be located at 1250 ng/J and 70% respectively.

Fig. 2 shows this kind of findings for burning raw and clean coals of both eastern and western products. Sulfur reduction through a physical coal cleaning unit, which is composed of dense medium vessel, dense medium cyclone and froth floatation. is assumed to be 32% for eastern coals and 17% for western coals. The percentage requirement for sulfur dioxide reduction is based on the sulfur mass content in raw coals. It should be mentioned that the majority of eastern coals are in a high sulfur category, therefore, their SO_2 removal requirement is always higher than that of western coals. Processes of the PCC unit will be discussed briefly later in the paper.

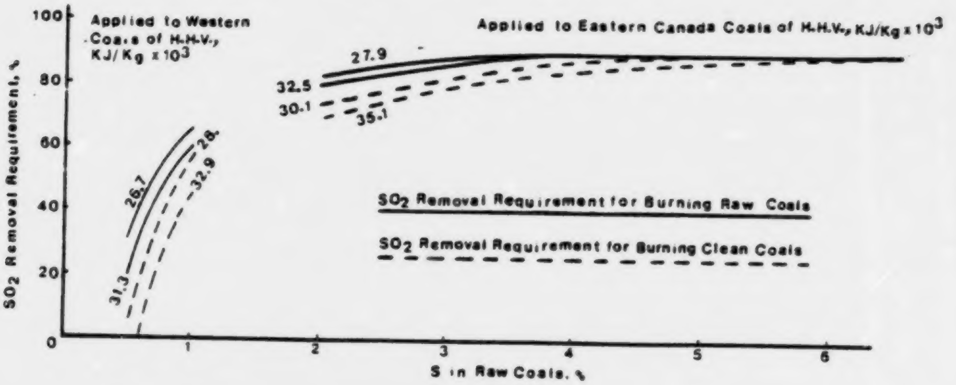


Fig 2 SO_2 Removal Requirement for Burning Raw & Clean Coals

As a result of stringent federal guideline and moderate efficiency of sulfur reduction in the PCC process, the requirement of SO_2 removal for burning eastern cleaned coals has no great difference from that for burning raw coals. The difference (see Fig. 2) becomes negligibly small as sulfur content in raw coals exceeded 5%, because, for coal H.H.V. ranged between 28000 and 32000 KJ/Kg, SO_2 emission level approaches or exceeds the limit level of 2580 ng/J. Over the limit, control regulation requires for 90% removal from the total emission.

ASPECTS OF PCC AND FGD TECHNOLOGIES

T. W. Tarkinto et al (3) studied three physical coal cleaning processes and three chemical coal cleaning processes. Among them a multi-stream physical process unit consisting of dense-medium vessel, dense-medium cyclone and froth floatation cells is a less expensive technology. The technically well developed and commercial available system is selected as a candidate for economical analysis in the present study. A popular process of limestone slurry scrubbing process is selected as a candidate in the FGD category. Their principal operations are briefly described below.

THE PCC PROCESS

The multi-stream process technology consists of three major unit operations. In the first stream, coarse coal, from raw coal preparation stage of crushing and screening, enters a dense-medium vessel for separation of clean coal and refuse, where coal rises on the upper level of the vessel and the refuse sinks toward the bottom.

The second one is an intermediate coal cleaning stream. Here a slurry of intermediate raw coal delivered from a fine screen in the preparation stage is pumped into a dense-medium cyclone via a feed sump for inertia separation of clean coal from refuse. Clean coal slurryes of the above two streams are dewatered in centrifuges before leaving the cleaning plant.

Fine raw coal slurry leaving from the screening section forms the third stream where fine clean coal is recovered from a series of froth floatation cells. After dewatering in a filter, the wet fine cleaned coal powder concurs with coarse and intermediate cleaned coals from the above and is ready for delivery to a stock pile.

The advantages of the multi-stream system are simplicity in dense-medium separating facilities and low cost. However, in order to meet with the stringent national guideline requirements, further treatment, in most cases as shown in Fig. 2, is still needed because high sulfur constituents remain in the cleaned coal.

THE FGD PROCESS

The FGD system selected here is a limestone slurry scrubber of turbulent contact absorber (TCA) type. The type of design has been one of the more successful mobile-bed absorbers applied to FGD, particularly in the mid-1970's. Recently, however, some competitive slurry scrubbers of mobile-bed designs, such as venturi-spray tower and spray tower, have been available for commercial applications. As far as costs are concerned, T.A. Burnett, el at (4) found the cost of TCA system is the lowest in capital investment and annual revenue requirements among the three absorber designs.

Basically, the TCA scrubber is a vertically countercurrent flow design where flue gas moves upward; and reagent slurry is sprayed toward the bottom of a vertical absorber. Before being forced into the absorber, flue gas is cooled and humidified by a spray of scrubbing slurry in a presaturator. Then it enters the main body of the scrubber, TCA, where four bed-layers of grids with adequate space apart are constructed in the absorber. On each of the grid bed foam spheres are packed, so that slurry and flue gas passing through each bed region will introduce a stream of turbulent flow to enhence slurry absorption efficiency and to prevent from scaling/plugging formation in the absorber. Leaving from the absorber, the gas continuously flows upward through a horizontal mist eliminator to minimize entrained moisture. In order to reduce corrosive action in the downstream and to increase buoyancy effect in the stack, scrubbed flue gas is heated above a saturated vapor temperature, say approximately 80° C, in a reheater where thermal energy will be consumed.

ECONOMIC EVALUATION

As shown in Fig. 2, SO_2 emission for burning PCC processed coals remains high and can not meet with the requirement of the guideline. Therefore the possible options to comply with the guideline will rely on (1) the PCC-FGD hybrid process and (2) FGD alone. Subsequent discussions will deal with the two categories.

COST ESTIMATES OF CAPITAL INVESTMENT AND ANNUAL OPERATING/MAINTENANCE (O/M) COST

Capital investment of major equipment of both PCC and FGD systems is extrapolated from the estimates in Ref. (3) with the following adjustments:

(1) a 10% escalation rate over the years of mid-1982 to mid-1985 (assuming the project beginning at mid-1983 and ending at mid-1985);

(2) 130% exchange rate for conversion of Canadian funds from U.S. funds;

(3) generating capacity designed at 1000 MWe in stead of 2000 MWe in the Ref. (3).

Indirect investments such as engineering design, construction expenses, contractor fees, contingency, etc. are estimated by making use of current Canadian market statistics. The equivalent present value of annual O/M cost is factored to account for 10%

annual discount rate and 3.5% annual inflation rate over an operation horizon of 30 years.

Coal consumption calculation is based on a 1000 MWe power plant operating at 10.19 x 10^3 KJ/KWh heat rate and 6000 hr/yr at MCR load. Coal prices are estimated to be 61 \$/10^3Kg and 55 \$/10^3Kg for eastern and western coals respectively. These prices are escalated from the statistics published in Canadian Minerals Yearbook, 1981 (5). It should be noted that all prices presented in the paper refer to Canadian funds.

In order to evaluate cost accounts within a wide sulfur content ranges for eastern and western coals as shown in Fig. 1, unit capital investment and unit O/M cost curves for both PCC and FGD systems are calculated in the present study. These unit price components refer to a unit removal rate of SO_2 in flue gas system for capital cost and to a unit mass SO_2 removal requirement for O/M cost. Figs 3 and 4 present the relationship of unit prices vs. sulfur content in raw coals under the reference conditions of 27840 KJ/Kg coal H.H.V. and 10189 KJ/KWh plant heat rate. If a situation differs from the reference conditions, a linear adjustment will be made on the cost analysis.

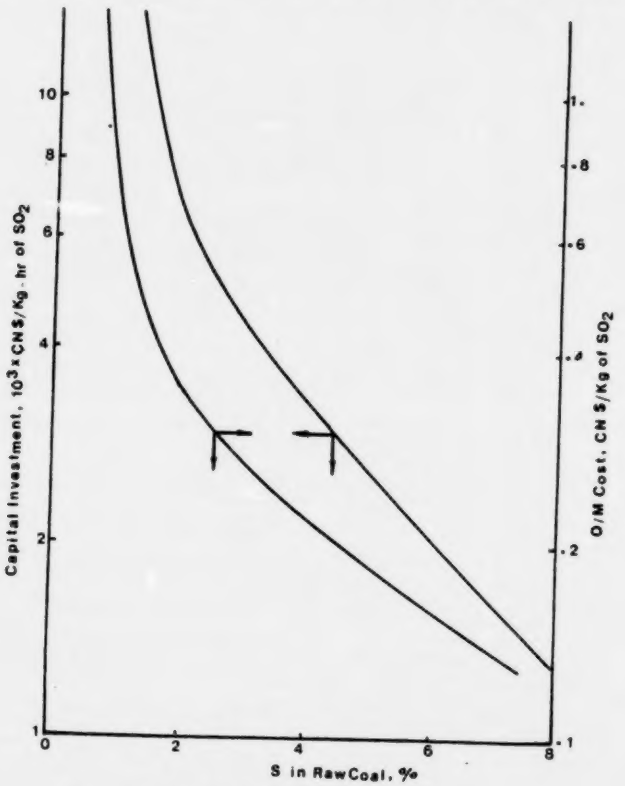


Fig 3 PCC Cost Estimate

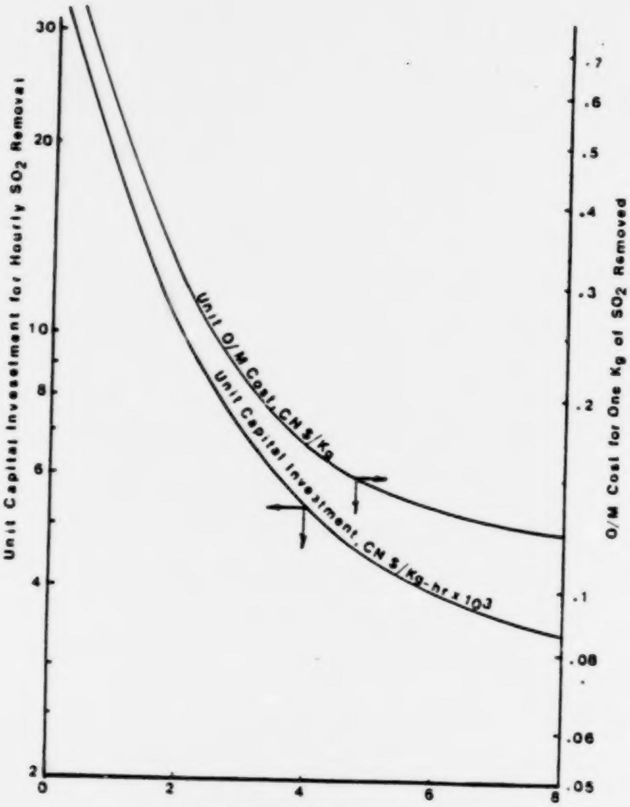


Fig. 4 FGD Cost Estimate

RESULTS OF COST BENEFIT ANALYSIS

In this paper study has been conducted on a comparison of life time cost between a PCC-FGD hybrid system and an individual FGD system in order to achieve a common goal of complying with a criterion defined by the national SO_2 emission guideline. For a given sulfur content in a coal, life time cost of the two systems reduces with increase of coal heating value. As a result, each of the two sulfur dioxide treatment systems will have a pair of scenarios, i.e. higher life time cost for poor heating value coal and vis-a-vis. Figs 5 and 6 reveal this type phenomenon.

In Fig. 5, analytical results show that the unit capital cost of PCC-FGD system is less expensive than that of the FGD system referring to eastern coals. But contrary situation occurs for western coals. Moderate cost difference in capital investment between the two systems appears in utilizing eastern coals. For western coals, the quantitative difference between the two systems becomes more vigorous. The results also show that unit capital cost of the FGD system increases with sulfur content whereas the cost of PCC-FGD system has a peak around 4% S content for eastern coals and around 0.75% for eastern coals.

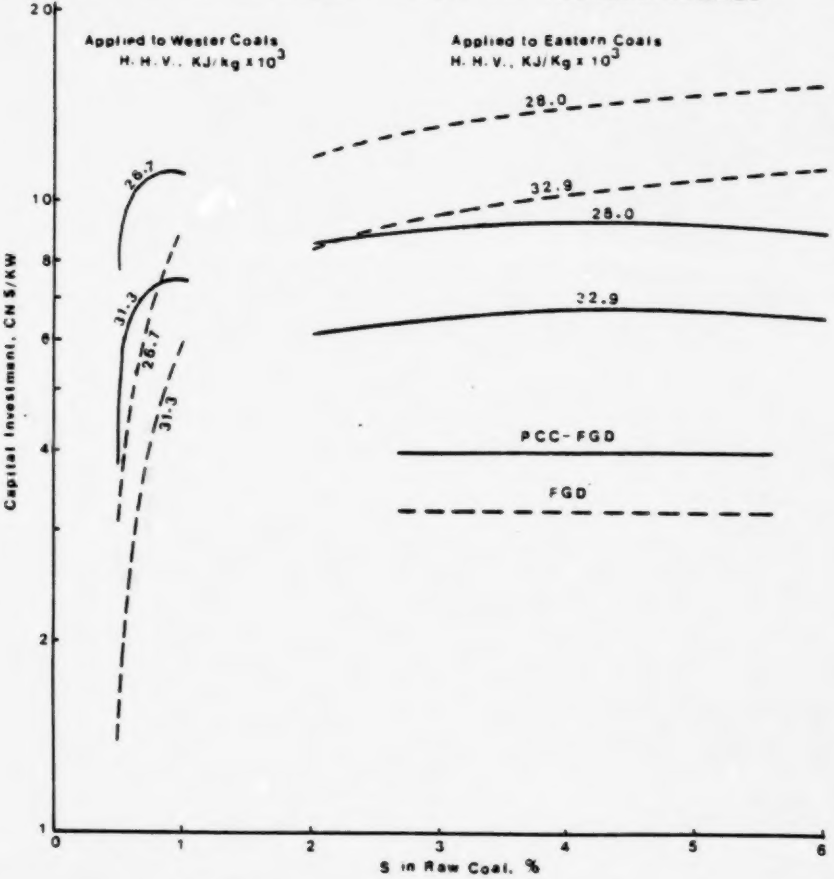


Fig 5 Capital Investment Cost Per KW Generating Capacity

In the present study, cost distribution indicates that capital investment in the total life time cost does not exceed 35% for the FGD system and 20% for the PCC-FGD system. Also, comparing unit O/M cost from Figs 3 and 4, we can see that for the same sulfur content the unit cost of the PCC system is always higher than that of the FGD system. High percentage O/M component in the total revenue requirement and high unit O/M for the PCC system lead the cost result of the PCC-FGD system into an inferior position in comparison with the equivalent unit revenue requirement for the FGD system. Fig.6 reveals this kind of tendency for eastern coals.

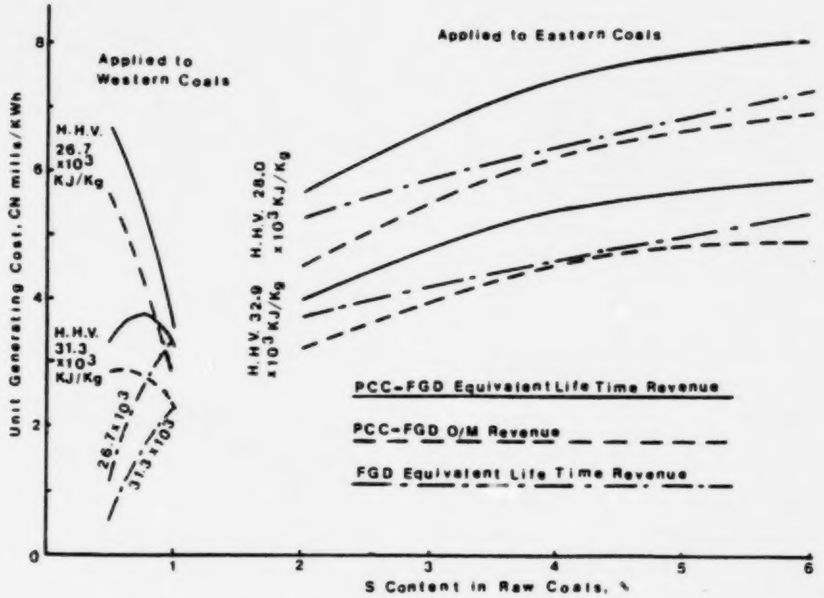


Fig 6 Unit Generating Cost of O/M Revenue and of Equivalent Life Time Revenue

In utilizing western coals, the dependence on FGD component in the PCC-FGD system becomes less important. Also, O/M cost for the PCC component drastically reduces with increase of sulfur content. These two factors give a marginal competition value to the PCC-FGD system around 1% S content region (see Fig. 6).

In conclusion, under the present preassumed conditions, the techno-economic analysis shows that due to the stringent SO_2 emission control guideline the PCC-FGD hybrid system is not favorable in competition with the individual FGD system. The situation may change, if the account of O/M cost for the PCC process can be improved in the technical and managemental aspects.

REFERENCES

1. Tibbetts, T.E. and Montgomery, W.J. "Evaluation of Canadian

Commercial Coals: Nova Scotia and New Brunswick - 1975" Energy, Mines and Resources Canada, Aug. 1976.

2. Tibbetts, T.E. and Montgomery, W.J. "Evaluation of Canadian Commercial Coals: Saskatwen, Alberta and British Columbia - 1975" Energy, Mines and Resources Canada, Aug. 1976.

3. Tankinton, T.W., Kennedy, F.M. and Patterson, J.G. "Evaluation of Physical/Chemical Coal Cleaning and Flue Gas Desulfurization" Technical Report EPA-600/7-79-250, Nov. 1979.

4. Burnett, T.A., Toritrick, R.L. and Sudhoff, F.A. "Economic Evaluation and comparison of Alternative Limestone Scrubbing Options" EPA/EPRI Symposium on FGD, Hollywood, FL. May 1982.

5. Canadian Minerals Year Book, Energy, Mines and Resources Canada, 1981.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

Size Measurements in Aerosols produced in Sulphur Dioxide Removal by Electron Beams

by

M.D. Carabine
P.G. Clay
G. Sisniega

Department of Chemical Engineering & Chemical Technology,
Imperial College,
London SW7 2BY
England.

Abstract

Droplet size distributions have been measured in the sulphuric acid aerosols which were produced when electron beams or plasma discharges were applied to gas streams containing sulphur dioxide.

For all the conditions investigated, the distributions were described by zeroth order log normal functions, and the modal diameters were in the range 0.3 to 0.5 µm. Half-peak widths were never greater than 0.3 µm., and the ZOLD spread parameters (σ_0) were in the range 0.1 to 0.5.

These sizes are judged suitable for effective removal by electrostatic precipitation; direct tests in this work confirmed this at least for very small (ca. 0.5 $dm^3.min.^{-1}$) volumetric flow rates. Flow rates about 100-fold higher were processed using the plasma method.

Electron beam currents between 10 and 100 µA produced conversions well above 90% with SO_2 concentrations in the range 100 to 5000 vpm, mixed in humidified air such that the molar ratio of oxygen to SO_2 equalled or exceeded 300. With flows up to 24 $dm^3.min.^{-1}$, 20 $mg.min.^{-1}$ of sulphur dioxide was removed.

INTRODUCTION

Laboratory experiments (1,2) and larger scale tests (3,4) have been reported in which electron beams have been utilised to remove sulphur dioxide from gas streams. Efficiencies of the process (5) and dependences on gas composition (6) have been investigated including the effects of additives such as ammonia. Our principal aims in the experiments reported here have been,

a) to investigate the conditions favourable for the SO_3 yield to take the form of a sulphuric acid aerosol, using a Van de Graaff accelerator.

b) to characterise the aerosols in respect of droplet size distribution, with a view to assessing their suitability for removal by means such as electrostatic precipitation.

c) to use a plasma discharge device, as an alternative to the electron beam accelerator, for promoting the oxidation of SO_2 to SO_3, and to characterise the consequent aerosols.

EXPERIMENTAL

Electron Beam Irradiations

Irradiations by electron beam were performed with a 2 MeV (High Voltage Engineering Corporation, Model AS 2000) Van de Graaff accelerator. The unit was fitted with a beam scanner which allowed the beam to be distributed along a target vessel of narrow cross-section, and length about 1.60m. Beam currents in the range 10-115 μA were used in these experiments.

The layout of the beam scanner, vessel, gas handling and analysis system are described elsewhere (7). Mixtures of SO_2 in N_2, or SO_2 in Ar, from cylinders supplied by BOC Special Gases, were mixed in a 5 dm^3 chamber with air previously saturated with water in three Dreschel bottles. The gas mixtures were passed under the beam in silica tubular vessels - typically with three passes and with tube diameters 0.5 or 1.0cm, and subsequently into a trapping solution of 0.1 mol dm^{-3} sodium hydroxide. The SO_2 absorbed was subsequently determined by iodimetric titration, and its conversion due to irradiation could be estimated by comparison with runs in the absence of the beam.

Success in producing aerosol was a primary objective of the experiments, and was readily detected visually in the trapping vessels. In suitable runs, a large plastic collector bag (capacity 70 dm^3; material polyethylene) was substituted for the absorber solutions, and used to transfer the aerosol to the test-space of a light-scattering apparatus for particle size analysis. The aerosols were expelled at constant flow rate into a laser beam (CW He-Ne) during scans of scattered light intensity at 15 angles in the range 9^0 to 130^0.

Interpretation of the scattered intensities was carried out by an inversion based on a zeroth-order logarithmic distribution of sizes (8). This method (9) produces modal diameters and spread parameters (σ_0) for distributions of spherical droplets which best fit the experimental data. The droplets were taken to be composed of sulphuric acid with a representative refractive index of 1.38 at 632.8nm.

Irradiations Using Air Plasma

The main features of the apparatus used with the plasma device are shown in Figure 1. The device itself has been described by F.J.Weinberg and co-authors (10). It has been shown to be capable of exceptional stability when the gas employed in the jet is N_2 or Ar; with air the running time is limited to some few minutes of continuous operation, due to problems with oxidation of the cathode material.

The device generates "pockets" of very hot gas which can survive for distances of some tens of millimetres downstream of the jet. Active atoms and radicals may diffuse from the "pockets" (11) into the cooler surrounding gas and undergo reactions there.

In the present arrangement (7,12), the SO_2-containing gas mixture to be processed was fed to the top of an outer cylindrical tube T1. The flowrates of these mixtures were in the range 90-300 $dm^3\ min^{-1}$. The plasma jet was supplied with air (or an inert carrier gas) at some $250cm^3\ s^{-1}$, and dissipated power at values between 300 and 1500 W. The flow from the plasma upwards into a central quartz tube T2 (1.5cm i.d. and 60 cm length) entrained the SO_2 air mixture, and the flow of products was split partly to a sampling bag (for droplet size analysis) and partly to exhaust. Temperatures were typically 110°C at the exit from the quartz tube, and 70°C in the collector bag.

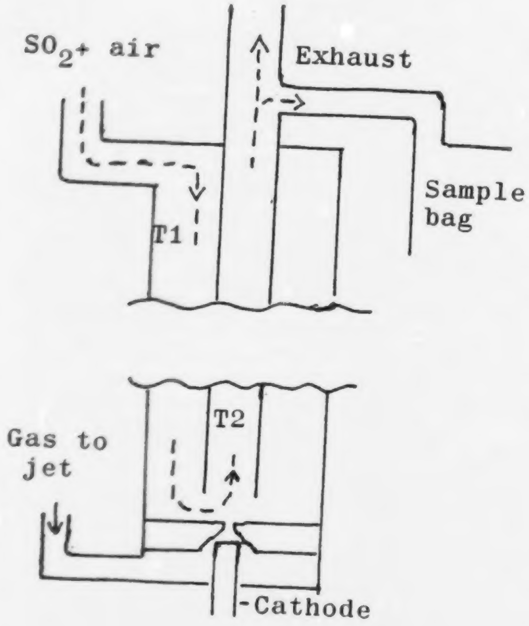


FIGURE 1. Schematic diagram of plasma-irradiation experiment.

RESULTS

1. Electron beam experiments.

a) Initially, tests with various gas mixtures and beam currents were made to identify conditions suitable for aerosol production

For irradiation in the absence of air, a typical result is shown in Figure 2, a mixture of 1000ppm SO_2 in N_2, with a beam of 100μA of 2 MeV electrons. The intercept on the titre axis corresponded to SO_2 collected in the "dead" time interval (ca.2 mins) between manually installing the absorbers (in the gas line close to the accelerator) and the first arrival of irradiated mixture in the absorbers.

Further results with N_2 and Ar carrier gases are in Table 1. None of these conditions produced detectable aerosol, but it was noted that substantial conversions of SO_2 resulted.

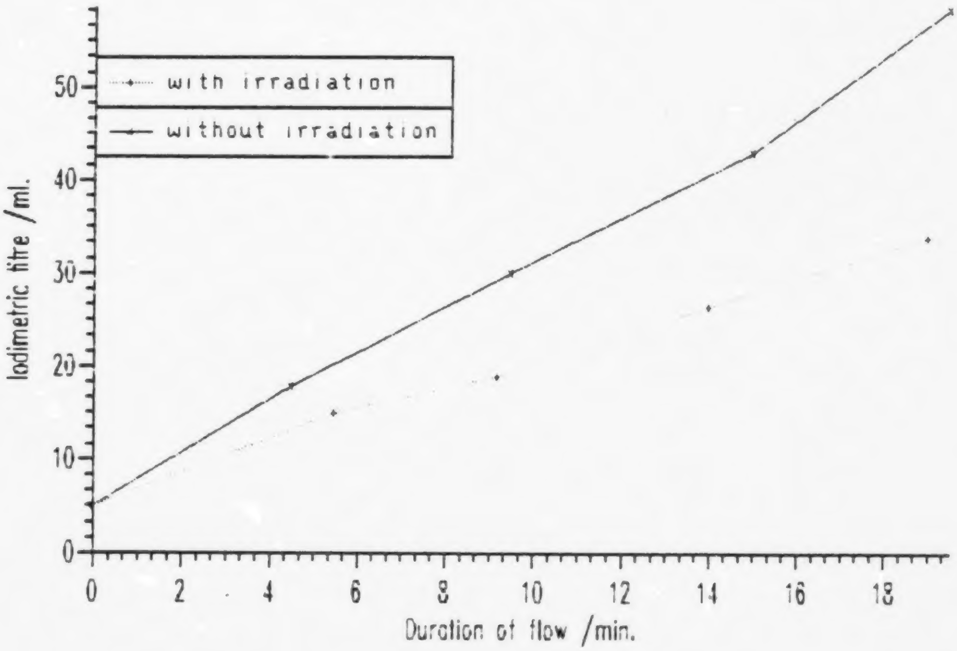


FIGURE 2.Removal of Sulphur dioxide in nitrogen by electron beam.

TABLE 1 CONVERSION OF SULPHUR DIOXIDE FOR ELECTRON BEAM IRRADIATIONS IN INERT CARRIER GASES.

Carrier gas	SO_2 concn./ppm	Beam Current/μA	Res.time/sec	Conversion
N_2	1000	100	19	40%
N_2	1000	100	16	60%
N_2	1000	25	14.5	15%
Ar	500	65	15	30%
Ar	125	65	5	31%

The effect of radiolysis in the presence of air, previously humidified, is typically of the form shown in Figure 3. In this case, a flow of 1000 ppm SO_2-N_2 mixture was mixed with a flow of moist air in respective amounts, 100cm^3 min^{-1} and 450cm^3 min^{-1}, resulting in a concentration of SO_2 before irradiation of some 180 ppm.

Titres are plotted against duration of run for beam currents of 10, 25, 30, 75 and 115μA. Conversions of SO_2 in the range 85-100% were recorded in all such cases, accompanied by satisfactory aerosol formation.

In contrast to these high conversions, a typical result with air taken from reagent grade cylinders, but not deliberately humidified, was about 60%, (in a mixture of 100cm^3 min^{-1} of 1000 ppm SO_2- N_2, with 500cm^3 min^{-1} of air; beam current 100μA). Aerosols were not visually detectable in these "dry" conditions.

The conversion of SO_2 in a given residence time plotted against the volumetric ratio of air to SO_2- N_2 mixture, is shown in Figure 4. The result suggests that a concentration of at least 300 moles of oxygen per mole SO_2 is necessary in these experiments for maximum conversion to be obtained.

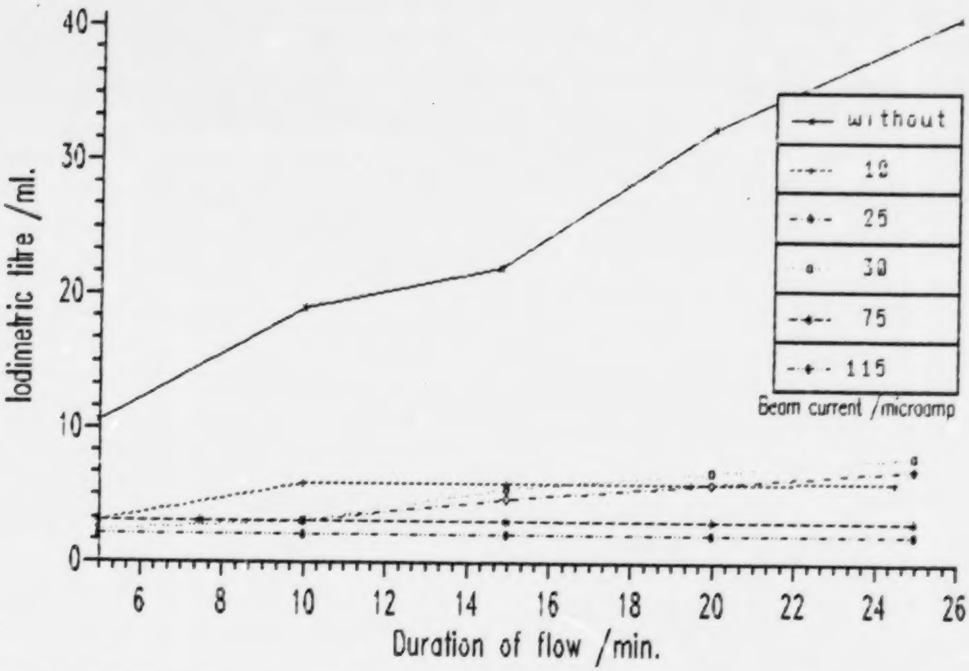


FIGURE 3. Removal of sulphur dioxide in humidified air by electron beam.

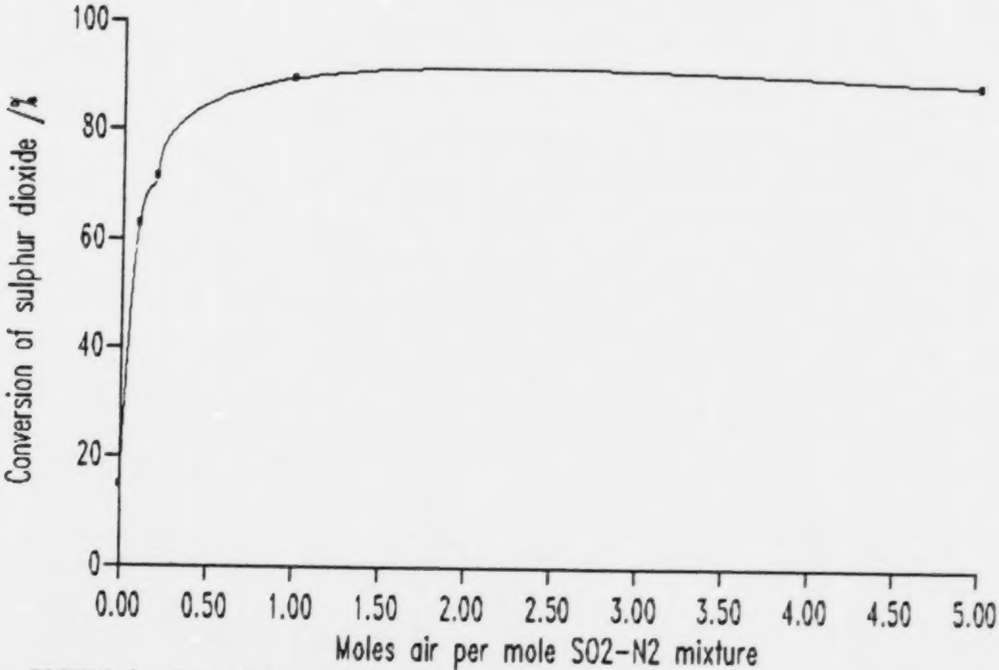


FIGURE 4. Dependence of removal of sulphur dioxide by electron beam on the proportion of oxygen in the mixtures.

b) Size distributions of aerosols from electron beam irradiation

Typical measured size distribution parameters are shown in Table 2. The reproducibility of results is demonstrated in Figure 5; aerosols produced in nominally identical runs showed modal diameters in the range 0.38 to 0.42 μm at a relatively constant value of σ_0. The standard procedure involved a delay of 5 ± 1 minutes between the end of collection at the accelerator, and the start of light scattering analysis; the negligible effect of the delay on the size distribution can be inferred from results such as those in Figure 6, comparing minimised delay with a deliberate interval of 45 minutes.

For concentrations of SO_2 in the range 10-56 ppm and reactor residence times of 3.5 and 12 sec., the size parameters were always in the range 0.3-0.6 μm for d_m, and 0.15-0.45 for σ_0. A trend of the distributions towards higher modal diameters with increasing SO_2 concentration is shown in Figure 7 (a) and (b).

The distributions in Figure 8 show a typical narrowing at 11.7 sec. residence time as compared with 3.6 sec; droplets of greater diameter than about 0.6 μm are apparently much fewer at the longer time. Visually the aerosols appeared more dilute, and absolute scattered intensities were significantly less at the longer residence times.

TABLE 2 PARTICLE SIZE DISTRIBUTION PARAMETERS FOR ELECTRON BEAM IRRADIATIONS IN AIR, RESIDENCE TIME 3.6s., (except "b" which is for 11.6s.).

Concentration of SO_2/ppm	10	13	15	18		29	56
				a	b		
d_m / µm	.42	.31	.33	.46	.41	.35	.48
σ_o	.25	.42	.18	.23	.20	.23	.26

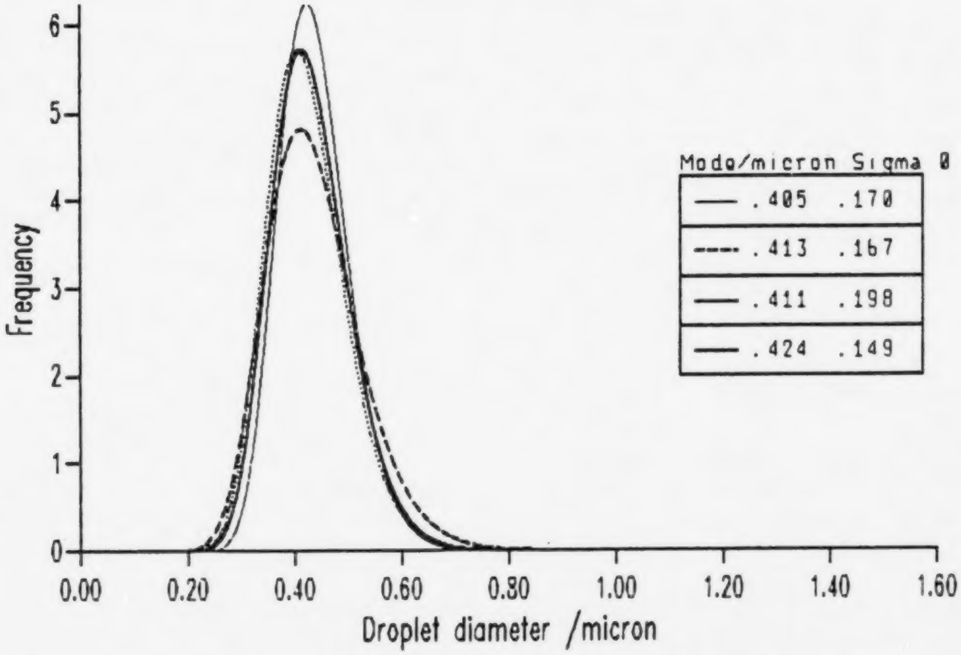


FIGURE 5. Reproducibility of particle size distribution measurement in successive runs.

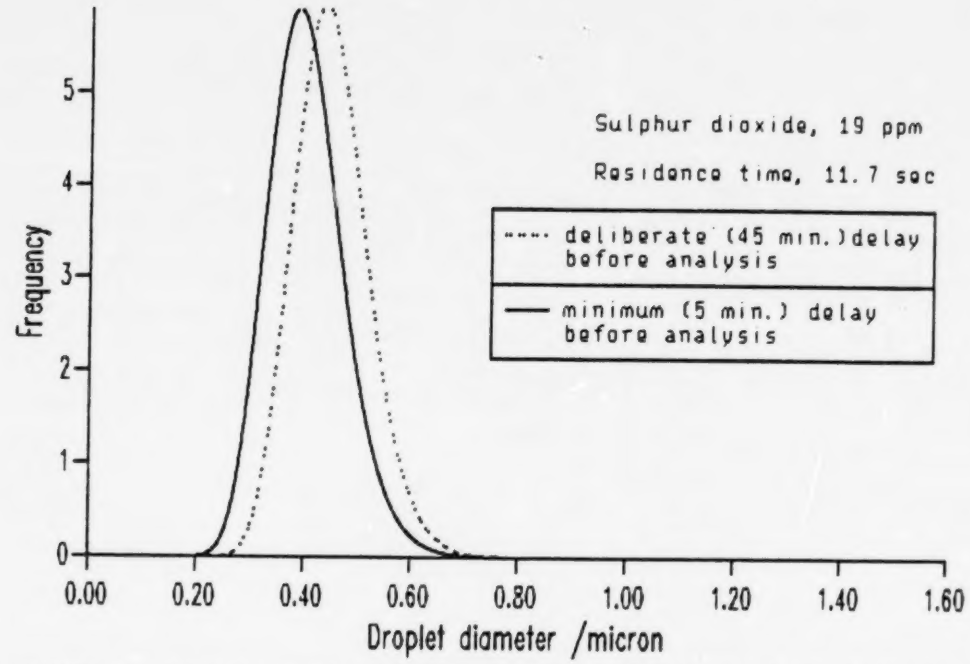


FIGURE 6. Comparison of particle size distributions after minimum and prolonged delays.

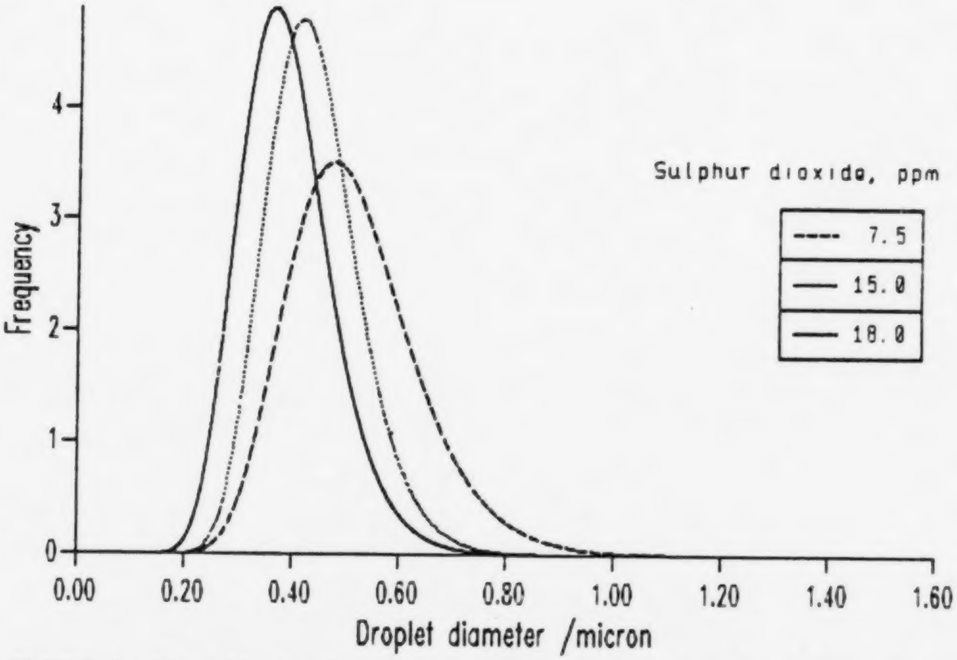


FIGURE 7. The effect of sulphur dioxide concentration on aerosol particle size distributions.

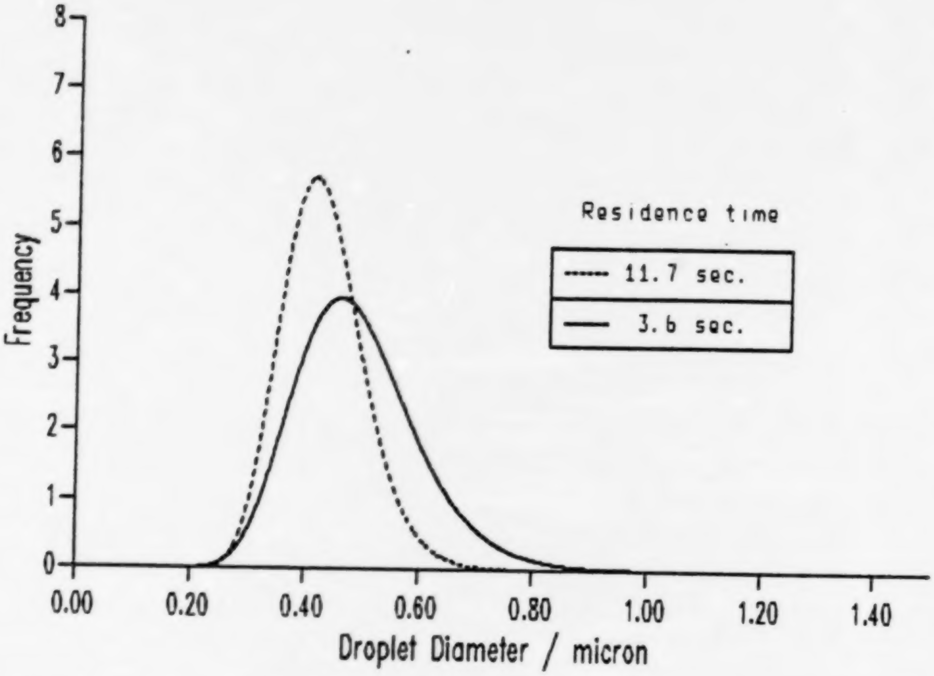


FIGURE 8. Typical variation of size distributions measured with different residence times under the electron beam.

c) Size distributions of aerosols from plasma-jet irradiation

Using the air plasma jet with inlet concentrations of SO_2 in the range 12-40 ppm, there was instantaneous production of sulphuric acid aerosol at high particle concentrations. There was also a high yield of nitrogen oxides under these conditions.

Argon-supported plasma which typically dissipated only 300-550 W produced no aerosol from SO_2 at similar concentrations and flow rates; runs with nitrogen-plasma resulted in no aerosol, except in a few as yet unverified cases. The particle size parameters obtained from analysis of air-plasma induced aerosols are shown in Table 3. The size distributions resemble closely those from electron beam irradiation, with the modal diameters in the range 0.3-0.5μm and σ_o in the range 0.1-0.45.

TABLE 3 PARTICLE SIZES OF AEROSOLS FROM PLASMA JET IRRADIATIONS.

Arc power /w	1600	1495	1495	1330	980
Air flow in plasma-jet/$cm^3 s^{-1}$	270	250	270	210	230
Air supply/$dm^3 min^{-1}$	140	150	150	80	270
Supply of 1000ppm SO_2-N_2/$cm^3 s^{-1}$	35	40	30	118	180
Concn. of SO_2 at plasma/ppm	13	14.3	11	71	37
d_m /μm	.39	.30	.41	.39	.45
σ_o	.45	.10	.12	.12	.28

DISCUSSION

There are a large number of possible reactions in irradiated SO_2-air-water mixtures but it is probable that the major steps will involve radiation products of N_2 and O_2, as the initial radiation energy deposition in these systems will be, to a first approximation, fractionated according to the volume composition. Despite the technological interest in these systems, rather little in the way of fundamental radiation chemical work appears to have been carried out. In consequence, a detailed discussion of the mechanism would be inappropriate. However, a fraction of the initial excited and ionised N_2 and O_2 molecules must ultimately appear as oxygen atoms, which are generally supposed (6) to be the actual oxidising species:

$$SO_2 + O = SO_3$$

The formation of SO_3 in mixtures of N_2-SO_2 containing no O_2 (Figure 2) but with some water content indicates that reactions between ionised and excited nitrogen molecules and water can also produce species capable of oxidising SO_2. Oxygen atoms could be again involved, but the production of hydroxyl radicals is also a possibility and these species are also known to be capable of oxidising SO_2 (13).

Figure 4 shows that increasing the oxygen content of the mixtures resulted in increases in the degree of oxidation of the SO_2 up to a point where there was at least 12% of oxygen in the system. Beyond this point, there appeared to be no dependence of conversion upon oxygen content. This type of behaviour occurs in many radiolytic systems and usually indicates the presence of competing reactions one or more of which leads to the product in question. In the present experiments, the competition presumably involved deactivation of the primary radiolytic species (e.g. excited or ionised nitrogen molecules) and reactions of these with oxygen to produce the oxidising oxygen atoms.

It is clear that the size distributions encountered in the aerosols produced by both methods of energy input - the electron beam and the plasma - suggest that removal of the sulphuric acid by electrostatic precipitation would be successful. It was possible to observe this qualitatively when the product from electron-beam irradiation was passed through a glass annular tube with D.C. voltage applied. Some inertial deposition from flows was also evident and might be exploited in the removal phase by suitable choice of residence times in scaled-up reactors.

Acknowledgements: We acknowledge important collaboration in the plasma technique by Dr. A.-M. Warris, and helpful encouragement of the project by Prof. F.J. Weinberg, F.R.S. Valued contributions to the experimental work were also made by S. N. Collins, A. L. Dyer, P. M. Harrison, and I. M. Waite. A studentship to G.S. by CONACYT of Mexico is acknowledged.

REFERENCES.

1. Kawamura, K. and Aoki,S. Study on application of ionizing radiation to flue gas desulfurization. Japan. Atomic Energy Soc. 14: 597, 1972

2. Kawamura, K., Aoki,S., Kawakami,S., Hashimoto,S. and Machi,S. Kinetic study of the irradiation of exhaust gases. I.A.E.A. Symposium, Vienna, 1975, p.633.

3. Kawamura, K. and Shui,V.H. Pilot plant experience in electron-beam treatment of iron-ore sintering flue gas, and its application to coal-boiler flue gas cleanup. Rad. Phys. and Chem. 24(1): 117, 1984.

4. Washino,M., Tokunaga,O., Hashimoto,S., Kawakami,W., Machi,S., Kawamura, K., and Aoki,S. Treatment of exhaust gases by irradiation. I.A.E.A. Symposium, Vienna, 1975, p.621.

5. Helfritch,D.J. and Feldman, P.L. Flue gas SO_2/NO_x control by combination of dry scrubber and electron beam. Rad. Phys. and Chem. 24(1): 129, 1984.

6. Tokunaga,O. and Suzuki,N. Radiation chemical reactions in NO_x and SO_2 removals from flue gas. Rad. Phys. and Chem. 24(1): 145, 1984.

7. Sisniega, G. Univ. London Ph.D. Thesis, 1983.

8. Espenscheid,W., Kerker,M. and Matijevic,E. Logarithmic distribution function for colloidal particles. J. Phys. Chem. 68: 3093, 1964.

9. Carabine, M.D., Maddock,J.L.M. and Moore, A.P. Technique for determination of particle size distributions in aerosols by light scattering measurements. 9th. Intl. Conf. on Atm. Aerosols Condensation and Ice Nuclei, Galway, Ireland, 1977.

10. Chan, A.K.F., Hilliard, J.C., Jones. A.R. and Weinberg,F.J. An electrically efficient finely tuneable low-power plasma generator J. Phys. D. Appl. Phys. 13: 2309, 1980.

11. Behbahani,H.F., Fontijn,A., Muller-Dethlefs,K. and Weinberg,F.J. Destruction of nitric oxide by nitrogen atoms from plasma jets. Comb. Sci. & Technology, 27: 123, 1982.

12. Warris, A.M. Univ. London Ph.D. Thesis, 1983.

THE NOXSO PROCESS DEVELOPMENT:

AN UPDATE

By:

J.L. Haslbeck, L.G. Neal

NOXSO Corporation
P.O. Box 586
Library, PA 15129-0586
(412) 675-6293

ABSTRACT

The NOXSO process is a dry, regenerable flue gas treatment system that reduces emissions of SO_2 and NO_x from coal-fired boilers by 90 percent. In 1982, NOXSO Corporation initiated a research and development program designed to develop the process for the commercial market.

This paper covers the work accomplished in the last six months. It is an update of two previous papers published in 1983 and 1984. The paper presents the results of research in the areas of process chemistry, sorbent performance, materials of construction and process design and economics.

DEVELOPMENT STATUS

A three-year test of the process at pre-pilot scale was completed in January 1985. The small-scale (50 ACFM), test facility was designed to obtain information on process chemistry and kinetics and sorbent life. Actual flue gas from TVA's Shawnee Unit No. 10, containing 2300 ppm SO_2 and 440 ppm NO_x, was used in the test. The key results of this test program were as follows [NOXSO Corporation (1983) and NOXSO Corporation (1984)]:

- Both sodium and alumina chemisorb SO_2 and NO_x from actual flue gas at 120°C. This means that less sorbent is required per mole of SO_2 and NO_x adsorbed than that predicted on the basis of sorbent sodium content alone.
- The sorbent's capacity is totally restored by heating to 600°C and treatment with a reducing gas. H_2, CO, H_2/CO, and H_2S have all been used to effectively regenerate the sorbent.
- A sorbent sample exposed to 45 cycles of adsorption and regeneration exhibited no deterioration in chemical activity toward SO_2 or NO_x.
- The rate of sorbent attrition has been shown in independent tests by NOXSO and PETC to be 0.02% of the fluidized bed inventory per hour. This rate does not change with increasing adsorption temperature (up to 300°C) or with the number of chemical cycles.
- A unique method of regenerating adsorbed NO_x has been proposed.
- The projected cost of a 500 MW(e) NOXSO flue gas treatment system has been developed by the Electric Power Research Institute (EPRI) using process design data supplied by NOXSO. The estimated capital and operating costs are 200 $/KW and 22 mills/KWh, respectively.

The success of the pre-pilot test has led to a pilot-scale demonstration of the process. On February 15, 1985, NOXSO and the Department of Energy signed a cooperative agreement to conduct a 1 MW(e) pilot test of the NOXSO process at PETC. Under the terms of this agreement, DOE will furnish all equipment and instrumentation needed to conduct the test. An existing test facility, that was used to evaluate the DOE fluidized bed copper oxide process, will be modified to conform to the NOXSO process design. The test will begin in May and run to the end of 1985. The goals of this test program are to:

- Determine the rate of sorbent attrition in a fluidized bed adsorber 40 times larger than the pre-pilot scale adsorber.
- Test the method of reducing NO_x by recirculating a concentrated stream of NO_x to the coal burner.

- Evaluate the effects of changing sorbent flow rate, inlet SO_2 and NO_x concentration, and bed temperature on the performance of the fluidized bed adsorber.
- Evaluate the performance and efficiency of the fluidized bed sorbent heater which is used to heat sorbent to 600°C prior to regeneration.
- Evaluate the effects of different reducing gases, changing inlet concentration of reducing gas, and changing gas flow rate on the performance of the sorbent regenerator.

The pilot test at PETC will also include a test on a smaller (300 SCFM), continuously operated system intended to produce a minimum of 300 cycles of adsorption and regeneration. This test system will be set-up so that all process steps are completely integrated. The test will generate important data on process integration and sorbent physical and chemical life.

PROCESS CHEMISTRY

The NOXSO process is a dry flue gas scrubbing system which employs a reusable sorbent. The sorbent simultaneously removes SO_2 and NO_x from flue gas at a temperature of 120°C. The spent sorbent is regenerated for reuse by treatment at high temperature (600°C) with a reducing gas. This regeneration reduces sorbed sulfur compounds to SO_2, H_2S, and elemental sulfur. The SO_2 and H_2S are then converted to elemental sulfur in a Claus type reactor. The sulfur produced is a marketable by-product of the process.

Adsorbed nitrogen compounds are decomposed and evolved on heating the sorbent to regeneration temperature. Regeneration of active NO_x sorption sites is, therefore, accomplished simply by heating the sorbent. The concentrated stream of NO_x produced on heat-up is returned to the boiler with the combustion air. This may be accomplished without increasing the concentration of NO_x in boiler flue gas because of the reversibility of NO_x formation in the boiler. At typical temperatures and gas residence times in the conventional pulverized coal-fired boiler, NO_x is in thermodynamic equilibrium with gaseous N_2 and O_2. The NO_x added to the combustion air either dissociates to N_2 and O_2 or depresses the formation of NO_x in the boiler so that the equilibrium NO_x concentration in the flue gas is maintained.

ADSORPTION

The sorbent is made by depositing Na_2CO_3 on the surface of an alumina substrate. Both the sorbent and the alumina substrate adsorb SO_2 and NO_x from the flue gas at 120°C. Figure 1 shows a series of identical sorption runs using NOXSO-1 sorbent (3.5% sodium) and alumina exposed to SO_2 in nitrogen and SO_2 in actual flue gas. The plot shows the adsorber effluent profile over time from a fixed bed reactor. The area above the curves is proportional to the quantity of SO_2 adsorbed. The activity of both the sorbent and the alumina is relatively low in the dry, synthetic gas. The

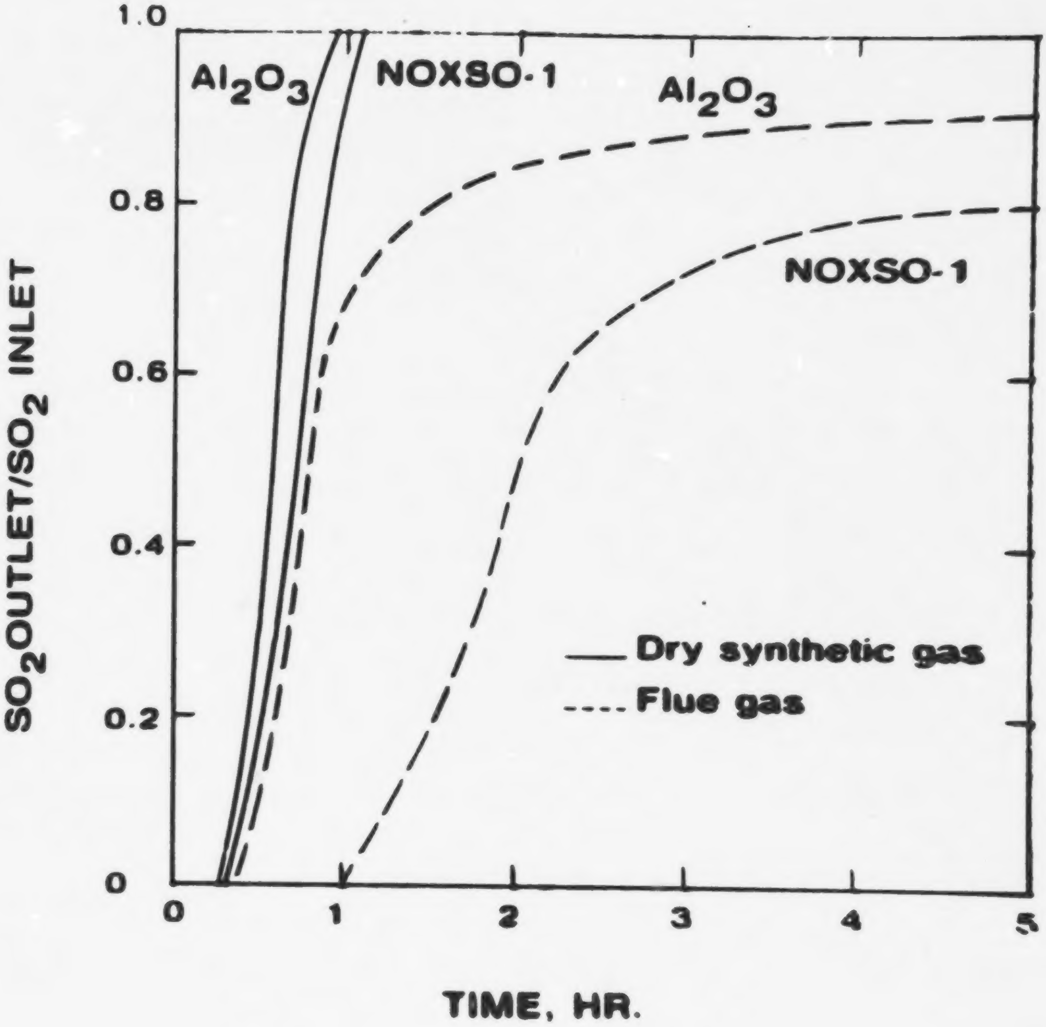


Figure 1. NOXSO Sorbent and Alumina Substrate in Actual Flue Gas and Dry Synthetic Gas.

quantity of SO_2 adsorbed increases by roughly a factor of 10 in actual flue gas. In the flue gas, the adsorptive capacity of the alumina is slightly less than half that of the sorbent containing 3.5% sodium.

The deposited Na_2CO_3 is unstable in the presence of alumina at between 400 and 700°C [Oden (1969)]. On the first pass through the sorbent regenerator, the sodium combines with the alumina to form $NaAlO_2$. Sodium aluminate adsorbs SO_2 as follows:

$$2NaAlO_{2(s)} + H_2O_{(g)} \rightleftarrows 2NaOH_{(s)} + Al_2O_{2(s)} \tag{1}$$

$$NaOH_{(s)} + SO_{2(g)} \rightarrow NaHSO_{3(s)} \tag{2}$$

$$NaHSO_{3(s)} + 1/2\ O_{2(g)} \xrightarrow{NO_x} NaHSO_{4(s)} \tag{3}$$

$$NaHSO_{4(s)} + NaOH_{(s)} \rightarrow Na_2SO_{4(s)} + H_2O_{(g)} \tag{4}$$

Steam and NO_x in the flue gas both play an important role in the SO_2 adsorption mechanism. Neither was present in the dry synthetic gas used to generate the data in Figure 1. By a mechanism similar to that shown above, sodium aluminate chemisorbs NO_x in flue gas to form $NaNO_2$ and $NaNO_3$.

The stoichiometry of reactions (1) through (4) is one mole SO_2 adsorbed per two moles sodium on the sorbent. Tests on NOXSO sorbent have shown the stoichiometry to be equivalent to one mole SO_2 per one mole sodium [NOXSO Corporation (1984)]. This is due to the fact that, in addition to sodium, alumina chemisorbs SO_2 from flue gas. The adsorption of SO_2 on alumina is well documented in the chemical literature [Deo (1971), Glass & Ross (1972), Khulbe & Mann (1978)]. The surface chemistry is complex and not completely understood. It is believed that chemisorption occurs on oxide ions (Bronsted bases) created by dehydration of the alumina surface. The adsorption products are sulfite and sulfate-like structures on the alumina surface. The products of SO_2 adsorption remain chemically stable as the sorbent is heated to the regeneration temperature of 600°C.

The alumina substrate will also adsorb NO_x from flue gas. Several researchers have demonstrated that NO_x chemisorbs on Lewis acid sites (aluminum ions) on gamma-alumina [Roev (1966), Pozdnyakov & Filimonov (1971), Turnham (1975)]. The product of NO_x chemisorption, in the case of both the NOXSO sorbent and the alumina substrate, is unstable at temperatures above 400°C. Therefore, the sorbent's activity toward NO_x is completely restored as the sorbent is heated to the sulfur regeneration temperature of 600°C.

REGENERATION

The regeneration of active SO_2 adsorption sites is accomplished by treating spent sorbent with a reducing gas. Hydrogen, CO, and H_2S have all been used to effectively regenerate the sorbent. The reduction of Na_2SO_4 with H_2 and CO proceeds as follows:

$$Na_2SO_{4(s)} + 4H_{2(g)} \rightarrow Na_2S_{(s)} + 4H_2O_{(g)} \tag{5}$$

$$Na_2SO_{4(s)} + Al_2O_{3(s)} + 4H_{2(g)} \rightarrow 2NaAlO_{2(s)} + H_2S_{(g)} + 3H_2O_{(g)} \quad (6)$$

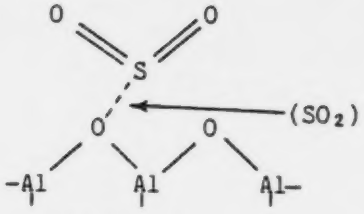
$$Na_2SO_{4(s)} + 4CO_{(g)} \rightarrow Na_2S_{(s)} + 4CO_{2(g)} \quad (7)$$

$$Na_2SO_{4(s)} + Al_2O_{3(s)}\ 4CO_{(g)} \rightarrow 2NaAlO_{2(s)} + COS_{(g)} + 3CO_{2(g)} \quad (8)$$

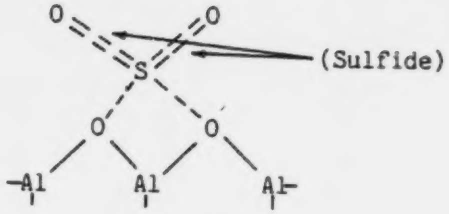
Tests have shown that less than 1% of the sorbed sulfur is converted to H_2S and COS via reactions (6) and (8) above. The sulfide produced via reactions (5) and (7) remains on the sorbent after treatment with the reducing gas. To fully restore the sorbent's activity, the sorbent is treated with steam to yield H_2S:

$$Na_2S_{(s)} + Al_2O_{3(s)} + H_2O_{(g)} \rightarrow 2NaAlO_{2(s)} + H_2S_{(g)} \quad (9)$$

The spent NOXSO sorbent contains alumina-sulfur compounds as well as sodium-sulfur compounds. Adsorption of SO_2 on alumina forms surface compounds of the type:



(Sulfite)



(Sulfate)

Regeneration of alumina-sulfur compounds yields a mixture of SO_2 and H_2S. It is believed that the regeneration product is determined by the point at which the reducing gas molecule severs the oxygen bond. Two possible reaction mechanisms leading to the formation of either SO_2 or the sulfide are illustrated schematically above. The alumina sulfide is readily hydrolyzed to H_2S by steam treatment.

Reduction of sorbed sulfur compounds yields a mixture of SO_2 and H_2S. The H_2S may be burned to produce a concentrated stream of SO_2, which then may be processed to produce sulfuric acid. Alternatively, the mixture of SO_2 and H_2S produced on regeneration may be converted to elemental sulfur via the Claus process. The typical distribution of products on regeneration of the sorbent and the alumina substrate with H_2 and CO is given in Table 1. The data show that, for both the sorbent and the alumina, at least two possible reaction mechanisms exist, leading to the formation of either SO_2 or H_2S. The data further show that H_2 and CO produce a significantly different mix of reaction products; H_2 producing predominantly SO_2 and CO producing predominantly H_2S. This difference could be caused by different orientations of H_2 and CO molecules on the alumina surface, since it is known that H_2 and CO attach to different types of active sites on alumina [Peri (1968), Bowen and Cheng (1974)].

The different mix of regeneration products formed by H_2 and CO makes it possible to adjust the proportion of SO_2 and H_2S produced on regeneration by adjusting the relative amounts of H_2 and CO used in the regeneration gas.

TABLE 1. Distribution of Regeneration Products
Alumina and NOXSO-1 Sorbent

Reducing Gas	Alumina H_2	Alumina CO	NOXSO Sorbent H_2	NOXSO Sorbent CO
Regeneration Products (Percent of Total SO_2 Adsorbed):				
SO_2, On Heating	15	12	10	11
SO_2, On Regeneration	58	38	72	25
H_2S, On Steam Treatment	11	41	15	64
TOTAL	84	91	97	100

A series of transient reaction tests were run to study the regeneration mechanism. In these tests, the reactant gas was abruptly changed at several crucial points in the regeneration of the sorbent. In a typical regeneration, SO_2 evolves immediately upon the addition of reducing gas to the fixed bed reactor. Following treatment with reducing gas, the sorbent is contacted with steam to convert sulfide to H_2S. In the transient study, steam was injected with hydrogen when the concentration of SO_2 in the reactor effluent reached its maximum. The rate of SO_2 evolution was not affected; however, H_2S evolved due to the reaction between steam and reduced sulfur species (sulfides) on the sorbent. The test showed that sulfides are formed throughout the regeneration process, including the period in which SO_2 evolution is at its peak.

The above test was repeated using CO as the reducing. Once again, the addition of steam produced H_2S but had no effect on the rate of SO_2 evolution. Following SO_2 evolution, the reducing gas was switched from CO to H_2. Hydrogen generates SO_2 more readily than CO, as shown in Table 1. However, the switch to H_2 at this point did not result in the formation of any additional SO_2, indicating that SO_2 is formed only in the initial stage of sorbent regeneration.

The results of these tests indicate that the reaction mechanism producing SO_2 is fast relative to the mechanism which produces the sulfide. The evolution of SO_2 occurs only in the initial stage of the regeneration process. The formation of SO_2 terminates before regeneration is complete, i.e., before the reduction of sorbed sulfur species has ended. The formation of the sulfide occurs throughout the regeneration process, even during the initial stage when the evolution of SO_2 is at its peak.

The product of NO_x chemisorption, in the case of both the sorbent and the alumina, is unstable at temperatures above 400°C. Therefore, the sorbent's activity toward NO_x is completely restored as the sorbent is heated to the sulfur regeneration temperature of 600°C. The concentrated stream of NO_x produced on heating the spent sorbent is recycled to the coal burner with the combustion air.

The NO_x returned to the coal burner does not increase the concentration of NO_x in the flue gas. The NO_x produced in the combustion process is in thermal equilibrium with N_2 and O_2 under the conditions that exist in the combustion chamber of a conventional pulverized coal-fired utility boiler. The NO_x returned to the boiler by the NOXSO process either suppresses the formation of NO_x in the boiler or dissociates to N_2 and O_2 which is released to the atmosphere.

A thermodynamic feasibility study of this method of NO_x reduction has been made using a Multophase Thermodynamics Equilibrium computer code developed in Pittsburgh Energy Technology Center, U.S. Department of Energy. The results show NOXSO's process for NO_x reduction is thermodynamically feasible.

Figure 2 shows equilibrium concentrations of NO at various flame temperatures. The thermodynamic model simulates pulverized coal combustion at 20% excess air and at different levels of heat losses. Recirculation of NO_x at 0%, 90%, and 100% was studied. The plotted results show the recirculated NO_x will be converted to N_2 and O_2 in the combustion chamber provided thermodynamic equilibrium is reached or closely approached [Yeh (1985)].

The NO_x reduction scheme will be demonstrated during the pilot test at PETC. NO_x will be added at varying rates to the combustion air entering the 500 lb/hr coal combustor at PETC. The burners used in the PETC coal combustor are similar to burners used in the utility industry. The concentration of NO_x in the flue gas will be continuously monitored to detect any change resulting from the practice of recycling NO_x back to the burner. This empirical evidence of the efficiency of the NOXSO NO_x reduction scheme will be available at the end of June.

SORBENT CHARACTERISTICS

Experiments were conducted to determine the long-term performance of the NOXSO sorbent. Two types of experiments were performed. The first experiment was a test in which a sorbent sample was fluidized for a long period in order to determine the physical attrition rate in the fluid bed. This test was a physical attrition test only. The fluidizing atmosphere was chemically inert.

The second set of experiments was a series of 45 cycles of adsorption and regeneration of a single sample of sorbent. The chemical activity of the sorbent toward SO_2 and NO_x was measured for each cycle and periodically the sorbent sample was removed from the reactor and submitted to an attrition test.

PHYSICAL ATTRITION RATE

A test was designed to determine the rate of sorbent attrition in the fluidized bed adsorber. This test was intended to measure physical attrition, i.e., attrition attributed to inter-particle collisions and col-

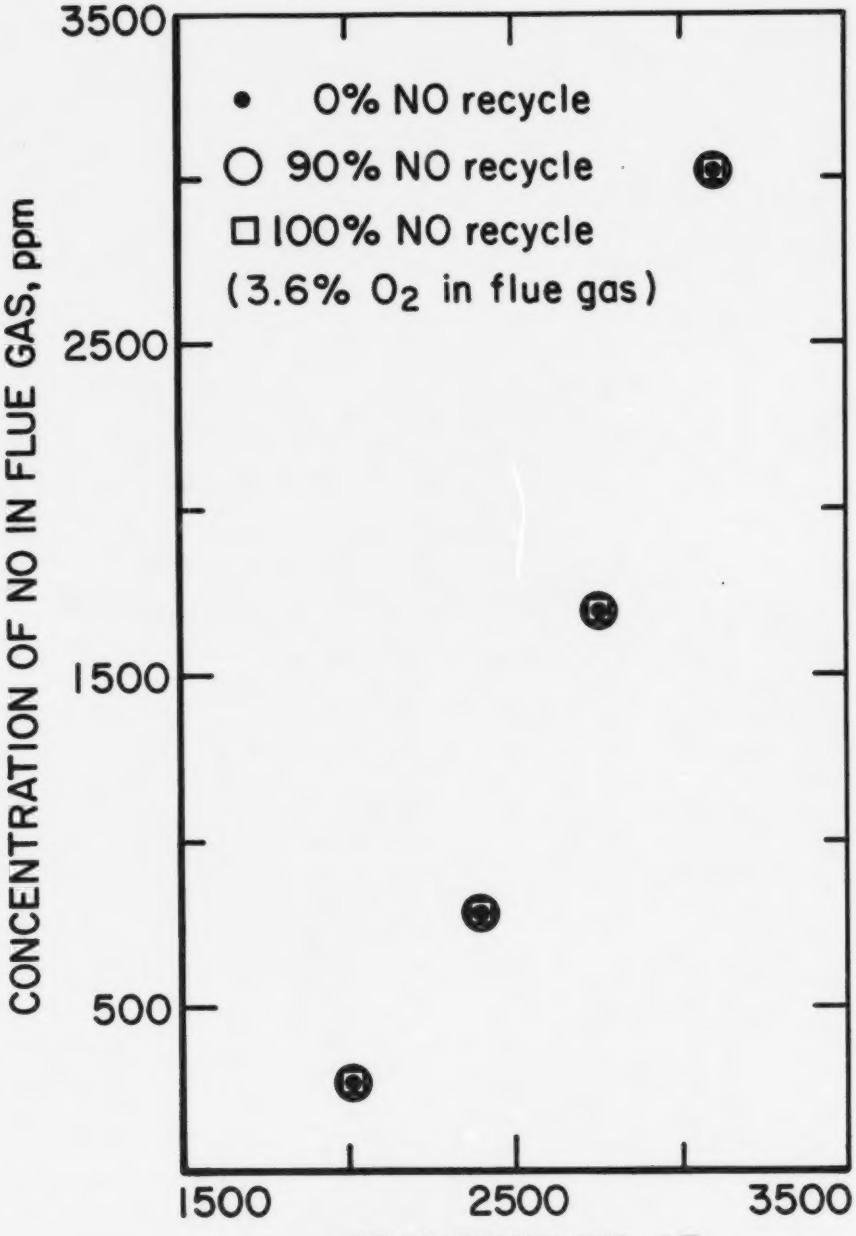


Fig. 2. Thermodynamic Equilibrium of NO in Flue Gas.

lisions with the walls of the reactor. The test did not include thermal or chemical cycling of the sorbent material.

The test apparatus was designed to simulate a NOXSO adsorber, as described in the section on process design. The key design factors are a bed depth of 18" and a superficial velocity of 2.4 ft/sec. The settled sorbent bed was supported on a 40 mesh screen fastened within a 3" diameter plexiglass pipe hung vertically. The pipe was unobstructed for six feet above the bed. The expanded (fluidized) bed height was 33 inches. The fluidizing medium was chemically inert, dry air. The apparatus held 1500 grams of NOXSO sorbent.

Table 2 shows the results of ten test periods totalling 576 hours (24 days). At each of the ten intervals, the sorbent sample was collected and weighed and then returned to the test apparatus with makeup of 1500 grams. The physical attrition rate is given in terms of the weight percent loss of bed material per hour. The rate of attrition over the ten test intervals varied from 0.015 to 0.025 %/hr.; the rate over the entire 576 hour test was 0.019 %/hr. This result agrees well with the independent tests conducted by DOE's Pittsburgh Energy Technology Center. DOE concluded that the NOXSO sorbent was similar to a sorbent material that exhibited an attrition rate of 0.02 to 0.03 %/hr. in tests conducted in a pilot-scale (40" by 48") fluidized bed adsorber.

TABLE 2. Attrition Rate Data

Test No.	Start Weight, gm	Final Weight, gm	Attrition Loss gm	Delta Time/ Total Time hr	Attrition Rate %/hr
1	1482.08	1458.30	23.78	67/67	.0239
2	1485.57	1470.60	14.97	45/112	.0224
3	1485.10	1473.10	12.00	40.5/152.5	.0200
4	1499.68	1485.08	14.60	66/218.5	.0148
5	1500.00	1494.85	5.15	22/240.5	.0156
6	1500.00	1475.07	24.93	89.5/330	.0186
7	1500.00	1490.44	9.56	40/370	.0159
8	1500.00	1488.80	11.20	46/416	.0162
9	1500.00	1479.21	20.79	67/483	.0207
10	1500.00	1474.50	25.49	93/576	.0183
			162.47	576	.0188

CHEMICAL ATTRITION

A sample of NOXSO-1 sorbent was subjected to 45 cycles of adsorption and regeneration. Adsorption of SO_2 and NO_x occurred at 120°C in actual flue gas from TVA's Shawnee Unit #10 containing, on average, 2300 ppm SO_2 and 440 ppm NO_x. Adsorption was continued until the concentration of SO_2 in the effluent from the fixed bed reactor reached 80% of the SO_2 inlet concentration, a period of 4 to 5 hours. Regeneration of the spent sorbent was accomplished with H_2 (30% by volume in N_2) at 600°C. Regeneration was terminated when the concentration of H_2 in the reactor effluent reached inlet concentration, a period of roughly 80 minutes. The sorbent was treated with steam immediately following treatment with H_2.

The sorbent was tested for attrition beginning at the fifth cycle and continuing at intervals of ten cycles for the duration of the endurance test. A 50 g sample was extracted following adsorption and subjected to an Accelerated Air Jet Apparatus (AAJA) Attrition test. The AAJA Apparatus and test procedures have been described in a previous report [NOXSO (1984)]. Following the attrition test, fresh sorbent was added to make-up to 100 g. The sorbent sample was then replaced in the reactor to continue cycling. At no time did the amount of fresh sorbent added as make-up exceed 6% of the total sample weight.

The results of the AAJA Attrition Tests are summarized in Figure 3. The bar chart shows the percent weight loss at 10 minute intervals during the 30 minute AAJA test. A test performed on fresh sorbent (no chemical cycling) is shown for comparison. The AAJA results after five cycles are comparable with the results of the test on fresh sorbent. At 15 and 25 cycles, the total weight loss during the test was less than half that of fresh sorbent. The weight loss was further reduced at 35 and 45 cycles. The results clearly show that the rate of sorbent attrition does not increase with chemical cycling.

The significant drop in weight loss after five cycles is attributed to the presence of "fines" in the sample at the start of the test. The fresh sorbent used in the endurance test was not subjected to an AAJA test at the start of the test. The results shown in Figure 3 for fresh sorbent were obtained from another sample of the same sorbent batch. The "fines" present in the original endurance test sample may account for a large portion of the weight loss measured after five cycles. The "fines", once removed, do not factor in the results of succeeding attrition tests.

SO_2 LOADING-ENDURANCE TEST

The quantity of SO_2 adsorbed was measured in 40 of the 45 cycles (for five cycles the recorder was being modified to accept thermocouple inputs and data were not taken). The results are plotted in Figure 4. The loading is calculated in units of cc SO_2 adsorbed (at 20°C, 1 atm.) per 100 grams sorbent. The data show a considerable amount of scatter ranging from roughly 2000 to 3500 cc SO_2 adsorbed (the first cycle of adsorption which is always higher than succeeding cycles is excluded from the analysis of the data in this section). Most probably, the scatter in the data is caused by incomplete regeneration of spent sorbent, which would cause the SO_2 loading

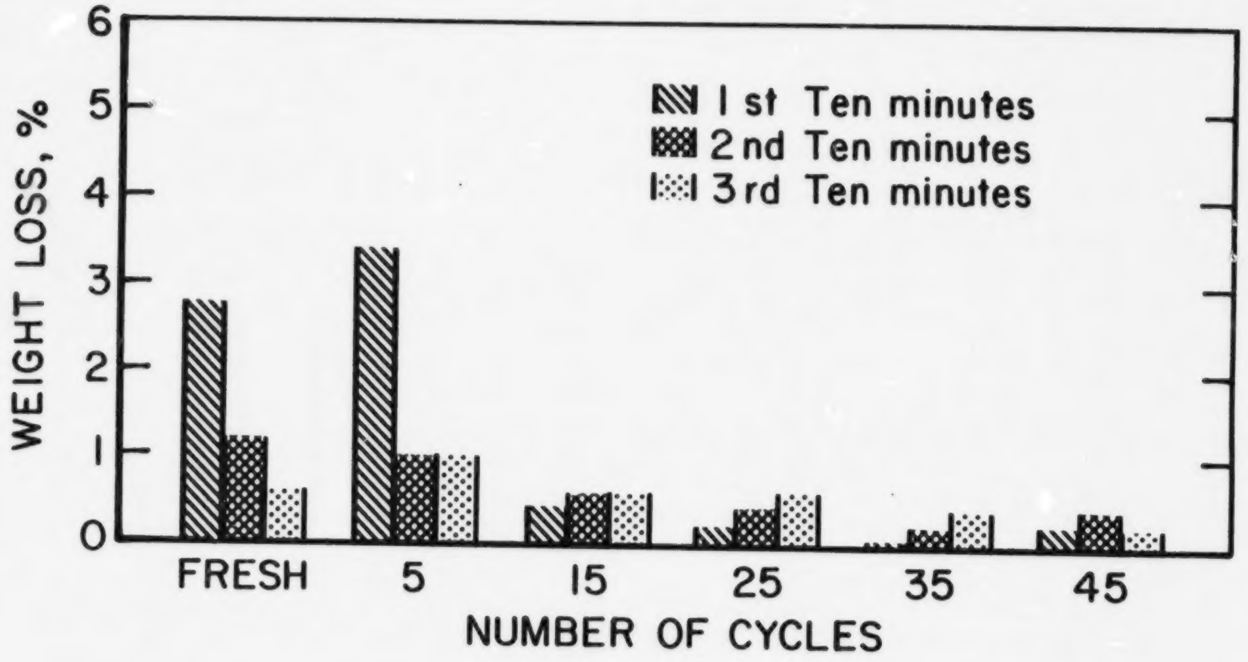


FIG. 3. ENDURANCE TEST — NOXSO-1 SORBENT RESULTS OF AAJA ATTRITION TESTS.

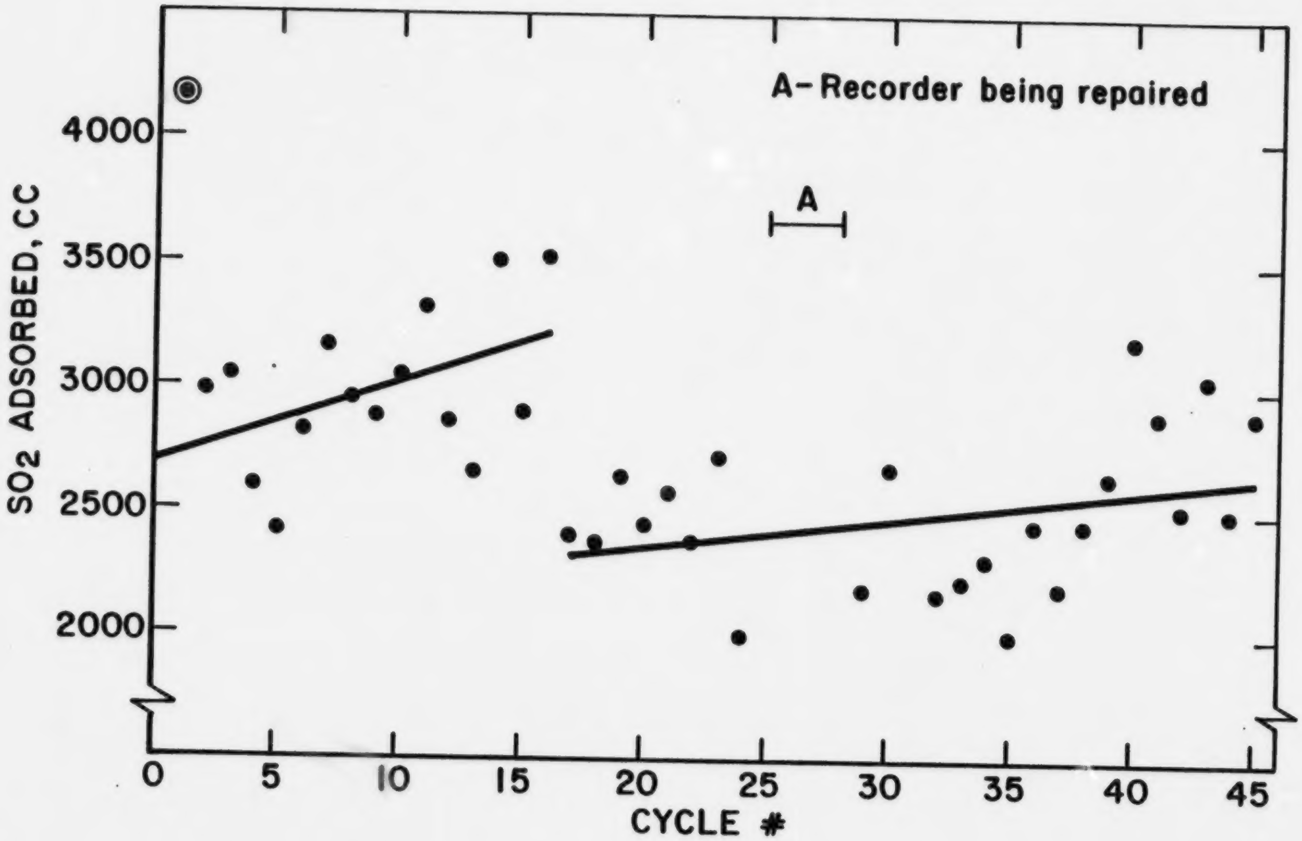


FIG. 4. ENDURANCE TEST — SO_2 LOADING NOXSO-1 SORBENT.

to be artifically low in the next adsorption cycle. Incomplete regeneration could be caused by an error in the regeneration procedure or by a progressive build-up of non-regenerable compounds on the sorbent. The former would produce a random scatter in the data; the latter would produce a discernable downward trend in the sorbent's performance.

The plot of Figure 4 shows a precipitous drop in SO_2 loading between cycles #16 and #17. This is believed to be caused by an error in the regeneration procedure. Normal procedure is to set the tube furnace to the desired temperature and to pass nitrogen through the reactor at 0.5 liters per minute until the bed temperature stabilizes at the regeneration temperature of 600°C. In cycle #17, the normal procedure was followed but the test apparatus was left unattended for a three hour period during heat-up. When the operator returned, the nitrogen flow to the reactor had stopped and the bed temperature had risen to 800°C.

There are several possible explanations for the reduction in sorbent performance at cycle #17. Exposure to high temperature could have produced compounds of higher molecular stability that were more difficult to reduce. The regeneration of cycle #16 might therefore be incomplete. High temperature could also calcine the sorbent more severely, driving off bonded OH groups thereby reducing the sorbent's activity in succeeding cycles. The calcining action might also be severe enough to convert gamma to alpha alumina which would reduce the sorbent's surface area and consequently its activity. The first two explanations are more probable since the sorbent's activity was restored after the high temperature episode. The conversion of gamma to alpha alumina, on the other hand, is known to be an irreversible process.

Due to the above, the data were divided into two groups: cycles (2-16) and cycles (17-45). The "best" straight line, using the least squares method, was drawn through the two sets of data as shown in Figure 4. In each case, the slope of the line is positive indicating that there is no progressive deterioration in sorbent performance, i.e., catalyst poisoning. The positive slope in the second group of data indicates that there was a gradual recovery of the sorption capacity that was lost due to operator error in cycle #16. The straight line fit shows a drop in sorption capacity of approximately 30% at cycle #17.

NO_x LOADING-ENDURANCE TEST

The NO_x loading was computed for roughly half of the cycles in which data were available. The results are plotted in Figure 5. The data scatter is evident, although not as pronounced as with the data on SO_2 loading. The NO_x loading ranges from a high of 493 cc to a low of 312 cc, with an average of 395 cc. There appears to be no drop in NO_x loading at cycle #17. This suggests that the drop in SO_2 loading noted above may indeed have been due to the formation of chemically stable sulfur compounds at high temperature, as opposed to the loss of surface OH groups which would be expected to affect both NO_x and SO_2 activity. All adsorbed NO_x was desorbed before the high temperature was reached.

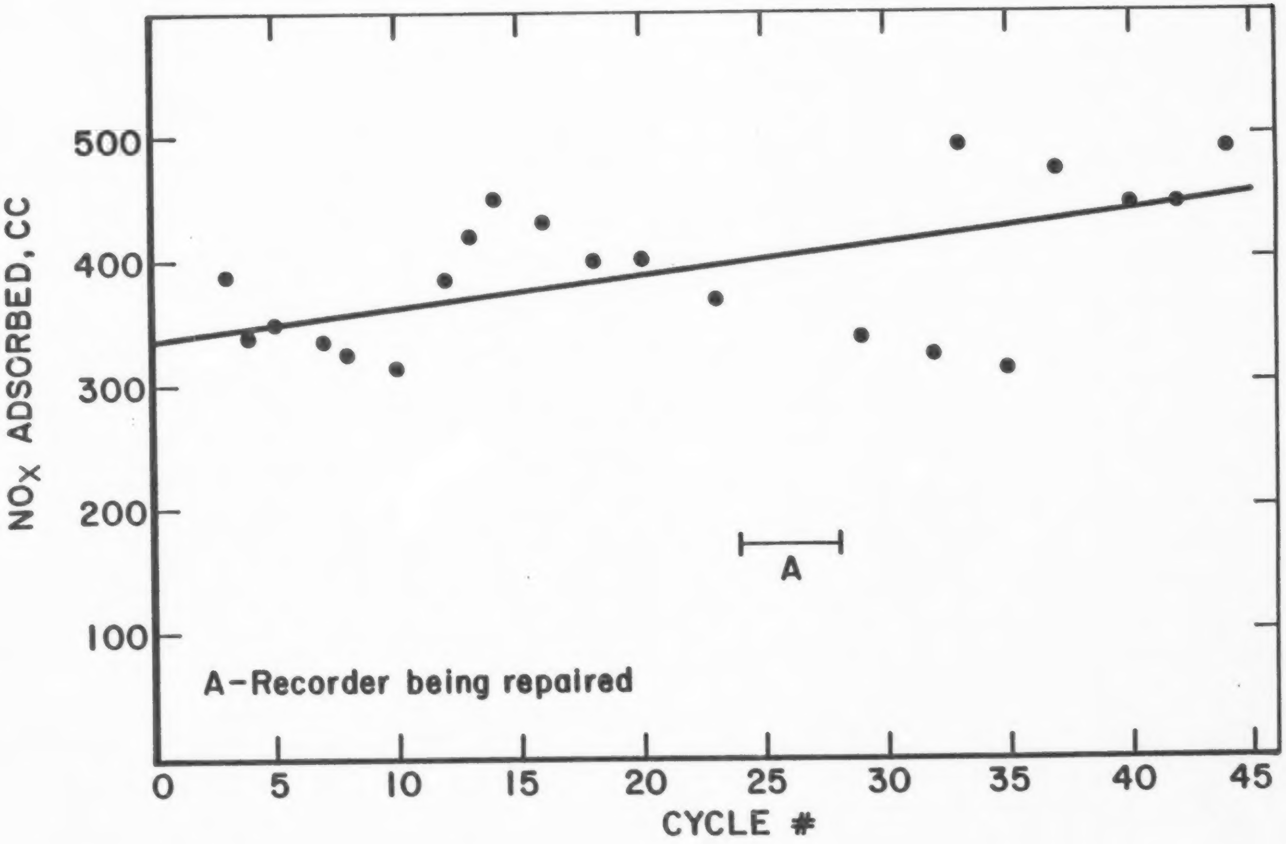


FIG. 5. ENDURANCE TEST — NO_x LOADING NOXSO-1 SORBENT

The best straight line through the data, using the least squares method, is shown in Figure 5. The slightly positive slope suggests that there was no progressive deterioration in NO_x activity over the 45 cycle test.

CORROSION AND ITS PREVENTION

In conducting the NOXSO test program, corrosion problems have been encountered in both the small scale reactor and the prepilot scale regenerator. A number of materials used in different parts of the system have failed or degraded. The materials used have included AISI 308, 316, 316L, and 446 stainless steels, Inconel 600, Monel 400, and Alonized (aluminized) 316L stainless steel. Inconel exhibited the most massive corrosion and AISI 446 stainless steel exhibited the least. The inside wall of the regenerator was made of AISI 446 stainless steel. A slight discoloration of the surface in contact with the sorbent was observed, but no scale was found in the vessel after approximately 200 hours of operation. All metals except 446 stainless steel and Alonized 316L showed scale. The corrosion results are summarized in Table 3.

TABLE 3. Corrosion Results of Some Selected Materials Used in the NOXSO Process

Parts	Materials	Results
Thermocouples	308, 316 316L stainless steels	Scale
	Inconel 600	Massive corrosion
Heaters	316, 316L	Massive corrosion confined to areas of sorbent contact
	Inconel 600, Monel 400	Massive corrosion confined to areas of sorbent contact
Small Scale Reactor	316, 316L	Scale
	"Alonized" Steel	Discoloration, but no scale
Large Scale	446 stainless steel	Discoloration, but no scale

Since only gaseous hydrogen sulfide (H_2S), sulfur dioxide (SO_2), sulfur vapor (S_x), and water vapor (H_2O), were involved in the regeneration reactions, one would expect that the corrosion is caused by high temperature oxidation or sulfidation. The temperature inside the vessels varied from 440-650°C, with most areas at 440-550°C. For the materials which have been used in this program, stainless steels are generally resistant to oxidation and sulfidation by these gaseous species in this temperature range. Ni-based alloys are resistant to oxidation, but easily attacked and embrittled by sulfur-bearing gases at elevated temperatures. If high temperature oxidation or sulfidation were the only cause of corrosion, one would not expect to see massive scale on AISI 316 or 316L stainless steels. Furthermore, any material susceptible to oxidation or sulfidation would have resulted in the most severe scaling at the hottest points. Yet, experimental evidence showed that the surface of metals in contact with sorbent suffered most severely. It was also found that thermocouples sheathed in stainless steels corroded only in the area contacted by sorbent. None or very little corrosion was observed in the area directly above the sorbent surface. In general, Ni-base alloys exhibited more severe corrosion than any other stainless steel materials.

Based on the test results, we believe that there is a corrosion type other than oxidation or sulfidation dominating in this program. The most probably one is "hot corrosion" caused by Na_2SO_4, deposited on the sorbent surface during adsorption, and in contact with the metallic parts during the regeneration reactions.

Hot corrosion was first observed in Ni-base superalloys used as turbine engine blade material at elevated temperatures. When Ni-based superalloys are used in gas turbine engines, the oxidation rates of these alloys are frequently much greater than those observed in pure oxygen or in air. It has been established that deposition of Na_2SO_4, resulting from salt ingested into the engine and sulfur from the combustion of fuel is a precursor to this type of accelerated oxidation or hot corrosion. The existance of sulfides in the corroded parts initially led to the erroneous conclusion that gaseous sulfur compounds are responsible for the accelerated attack. However, the oxidation of presulfidized alloys is generally not accelerated, and shows different oxidation behavior [Goebel (1970)]. The attack of gas turbine alloys by fused Na_2SO_4 is characterized by the presence of a loose, non-adherent oxide scale that is separated from the unaffected base material by another layer of material which contains sulfur rich precipitates. Some scales formed in our test program followed this characteristic pattern.

NOXSO Corporation conducted an extensive study of materials and their application in the NOXSO process. It was concluded that austenitic stainless steels may be subject to hot corrosion from Na_2SO_4, by sulfide formation and to a limited extent from basic fluxing due to the Ni content. Stainless steel (type 316) is subject to acidic fluxing induced as a result of its Mo content. Type 446 has better corrosion resistance due to the higher Cr content and lack of Ni.

Several stainless steel and Ni-base alloys used as sheath or shielding materials for thermocouples, heaters, and reactors have failed or degraded. Ni-base alloys (Inconel, Monel) exhibited the worst corrosion, 308, 316, and 316L stainless steel were slightly less corroded. Alonized steels and type 446 stainless steel performed satisfactorily. Based upon the previous discussion of high temperature oxidation, sulfidation, and Na_2SO_4-induced hot corrosion, conclusions are drawn as follows:

1. Oxidation is generally not a problem except that sensitization may occur in austenitic stainless steels, causing serious intergranular corrosion.

2. Ni-base alloys are incompatible with sulfur-bearing gases and especially susceptible to hot corrosion attack. Formation of liquid nickel sulfide phases can even severly disintegrate the complete metal structure.

3. Austenitic stainless steels are fairly resistant to sulfidation in the NOXSO System service temperature range, but the deleterious Mo content in type 316 (or 316L) may cause severe problems by hot corrosion.

4. Type 446 ferritic stainless steel seems satisfactory in all environments experienced because of its high Cr content and lack of Ni.

In general, Fe-base alloys are more resistant to sulfur attack and Na_2SO_4-induced hot corrosion than Ni-base alloys. Since in the reactor the metallic parts are in contact with the sorbent which is covered with Na_2SO_4 and the regeneration process involves gases S_x, H_2, SO_2, and H_2O, Ni-based alloys should not be considered as candidate materials in this system. Type 446 ferritic stainless steel may be a good base material except that it is relatively more expensive than austenitic stainless steels. Austenitic stainless steels with low carbon and without Mo content (e.g., 304L) can be satisfactory if a surface coating is applied to improve the hot corrosion resistance. From an economic point of view, even cheap low alloy steels may be considered provided they are well treated by aluminization or other coatings.

Currently, materials used in high temperature, high performance systems (superalloy, steel, refractory steel, and ceramic components) are often subjected to surface treatments with aluminization or MCrAlY overlay coatings to improve the resistance to high temperature oxidation, sulfidation or hot corrosion. These types of coatings offer the best resistance to scaling in sulfur-oxygen gases. The general principles and applications of aluminization and MCrAlY overlay coatings are briefly discussed here. More detailed information on the performance of these high temperature, protective coatings are given in a recent TMS-AIME symposium [Singhal (1982)].

The "Alonized" (aluminized) 316L stainless steel used as the inside wall material of the reactor exhibited good resistance to high temperature corrosion in the NOXSO test program. "Alonized" steels were coated by a pack cementation process by Alon Processing, Inc., Tarentum, Pa. This treatment, commonly referred to as "alonizing", diffuses aluminum into the metal surface to form a high-melting aluminum intermetallic. A two-layer coating structure is normally observed. The outer layer has a relatively high Al content (AlFe intermetallic) whereas the inner protective layer has a somewhat lower content of Al and Fe and appears to be a solid solution alloy of Al and Fe. This aluminide coating has been widely used to improve the high temperature corrosion resistance of Ni, Co, and Fe-base alloys. Aluminizing offers excellent performance under oxidation conditions up to

1100°C. With S and Na present, aluminizing improves material performance in the 800-900°C temperature range.

Aluminizing can be applied either by a pack process or a vapor deposition process. In the pack method, the coating is formed by treating parts at a high temperature in a pack box, which contains a mixture of Al powder, an inert filler material, Al_2O_3, and a chloride activator, NH_4Cl. A typical composition can be $72Al_2O_3$-25Al-$3NH_4Cl$. The coating obtained consists of AlFe (or NiAl) layer which forms an aluminum oxide scale on its surface during oxidation. In the vapor phase aluminizing (either by electron beam, physical, or chemical vapor deposition), aluminum is transferred from the pack to the substrate through the gaseous phase. This requires the simultaneous presence of an Al compound (e.g. 70Cr-30Al) and of a halide (e.g. AlF_3) which allows the Al transfer. The deposited Al reacts with the substrate to form aluminides at high temperatures. Another method is plasma spraying, where an Al-rich powder is plasma sprayed onto the substrate and electron beam remelting of the sprayed coatings is then carried out to obtain complete fusion of the coating. A coating thickness of 25 to 125 μm is usually sufficient for surface protection.

In an oxidizing environment, particularly at high temperatures, aluminides are preferred, whereas at moderate temperatures and in sulfidizing conditions, chromizing exhibits better performance. Chromizing can be applied by a pack process similar to aluminizing except that in this case, pure Cr is used instead of a mixture of compounds. Chromizing followed by aluminizing or co-deposit of Cr and Al allows combining properties of both coatings to meet the characteristics required for the system.

A more sophisticated coating produces an overlay coating of MCrAlY (where M = Fe, Co, or Ni). This may be produced by condensing a multi-element vapor on the metal surface. The coating contains Cr and Al for corrosion resistance where Cr ranges from 18 to 35 wt%, and Al from 6 to 13 wt%. A typical composition of the coating is 21% Cr, 11% Al, 0.3% Y and the balance M. Continuous layers of Al_2O_3 and Cr_2O_3 are formed on these alloys when they are exposed to gases containing oxygen. Yttrium is used to improve the adherence of the oxides to the alloy. Selection of M (Fe, Co, Ni or a combination of these elements) depends upon the intended use of the coatings. For example, Co, 18-22% Cr, 10-12% Al, and 0.2-0.5% Y has been used to protect alloys from hot corrosion attack effectively to 900°C, but substantial attack was observed at T = 650-800°C. Fe is better than Ni and Co for T = 800°C.

Some more resistant, yet more expensive, ceramic thermal barrier coatings have been developed for aircraft and industrial gas turbine operations. These coatings consist of a layer of a plasma-sprayed ceramic oxide applied over a plasma-sprayed metallic bond coat. Ceramic materials such as zirconia-yttria (ZrO_2-Y_2O_3) and calcium silicate (Ca_2SiO_4) are commonly used for the outer layer and the bond coat material is usually of the MCrAlY type. Graded coatings consisting of an MCrAlY bond coat/MCrAlY-oxide graded zone/oxide over coat may be even better to minimize the thermal expansion mismatch between the ceramic coating and the metal substrate.

The high temperature corrosion behavior (oxidation, sulfidation, and hot corrosion) of a number of Ni-based and Fe-base alloys has been studied. Ni-base alloys are found to be incompatible with the compounds involved in the NOXSO Process regeneration reactions. Austenitic stainless steels are susceptible to sensitization and to a limited extent, hot corrosion. Type 446 ferritic stainless steel with high Cr content and no Ni performs satisfactorily under the service conditions. Improvement of high temperature corrosion resistance can be achieved by surface coatings such as aluminizing, chromizing, McrAlY overlay coatings, or ceramic coatings. Aluminized 316L stainless steel has proven effective against the corrosion attack. Other base materials such as low alloy steels, low Cr-content ferritic stainless steels, or low carbon austenitic stainless steels may also be considered, provided good surface coatings are applied.

PROCESS DESIGN AND ECONOMICS

The NOXSO system configuration has evolved over three years of development. As each series of tests are completed and the data analyzed, the process design is reviewed and modified. The continuing goal is to optimize the system to obtain low cost, both capital and operating, and simplicity of operation. A number of changes in the design have been made as a result of earlier cost studies.

PROCESS DESIGN

The current design concept is illustrated in the flow sheet of Figure 6. The process operation is described by following, in turn, the flow of flue gas, sorbent, regenerant gas and combustion air through the system. Flue gas leaving the coal-fired boiler passes through the combustion air pre-heater (Unit #2) into the flue gas treatment system. It first enters a temperature control system (Unit #3) in which water is sprayed into the flue gas to reduce its temperature to about 200°F by vaporization. The flue gas then passes through the booster fan (Unit #4), and enters the fluid bed adsorber (Unit #5). Note that the flue gas still contains the full load of fly ash. The fly ash is typically 2 to 3 orders of magnitude smaller than the sorbent particles and will pass directly through the fluid bed adsorber. The gas exits the adsorber into the main flue gas line, subsequently passing through a electrostatic precipitator (Unit #6), a flue gas fan and is released to the atmosphere through the stack. Just prior to entering the ESP, SO_3 is added, as necessary, to improve the performance of the ESP. The amount of sulfur added as SO_3 is about 1% of the total sulfur removed in the adsorber.

Within the adsorber, the NO_x and SO_2 are simultaneously removed from the flue gas stream. Spent sorbent is continuously removed from the fluidized bed adsorber and transported to a three-stage fluid bed heater (Unit #9). Make-up sorbent is added to the sorbent stream as it exits the adsorber and before entering the fluid bed heater (Unit #11). This fresh sorbent increases the sorption capacity of the spent sorbent stream and allows for removal of small quantities of sulfur dioxide contained in the heating gases. In the sorbent heater, the sorbent is passed counter-current to a stream of high temperature gas which heats the sorbent to the regeneration temperature. The NO_x and a small amount of loosely held SO_2 are

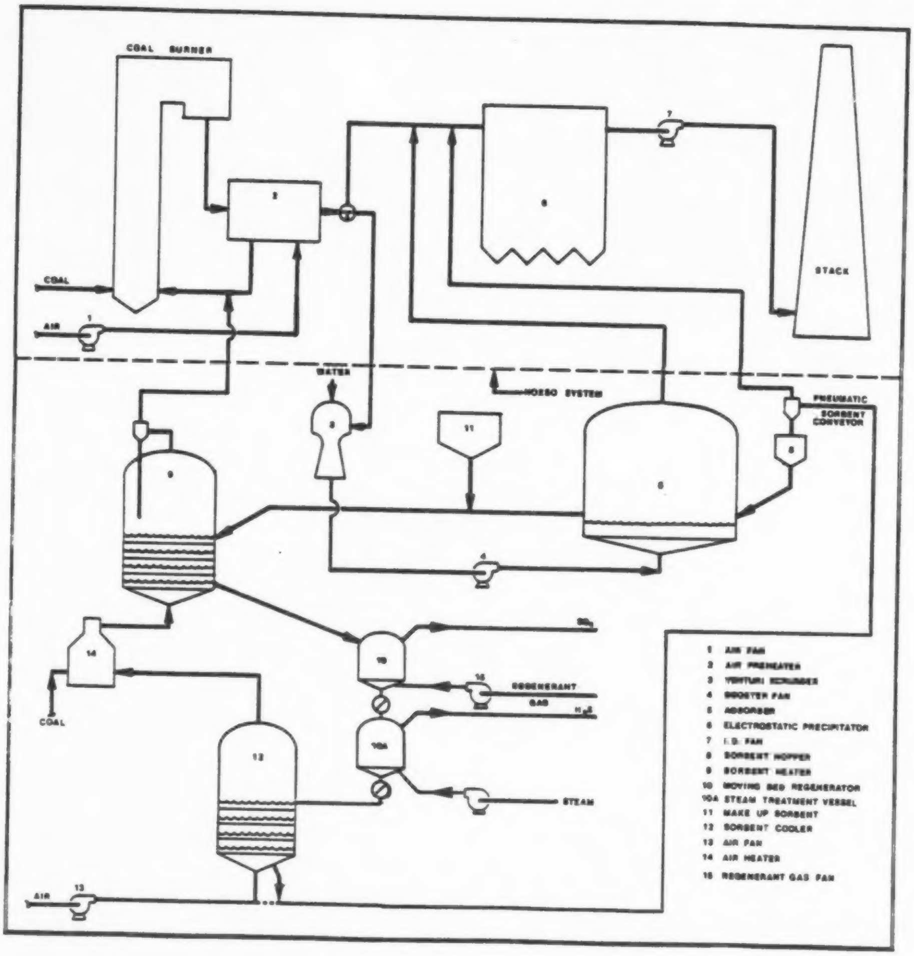

COAL BURNER
2
6
7
STACK
COAL
AIR
1
WATER
NOXSO SYSTEM
PNEUMATIC SORBENT CONVEYOR
3
11
9
5
8
4
SO2
10
15
REGENERANT GAS
H2S
10A
STEAM
14
COAL
12
AIR
13
1 AIR FAN
2 AIR PREHEATER
3 VENTURI SCRUBBER
4 BOOSTER FAN
5 ADSORBER
6 ELECTROSTATIC PRECIPITATOR
7 I.D. FAN
8 SORBENT HOPPER
9 SORBENT HEATER
10 MOVING BED REGENERATOR
10A STEAM TREATMENT VESSEL
11 MAKE UP SORBENT
12 SORBENT COOLER
13 AIR FAN
14 AIR HEATER
15 REGENERANT GAS FAN

Figure 6. NOXSO Flue Gas Treatment Process Flow Sheet.

stripped from the sorbent in this heating operation and carried away in the effluent. The heated sorbent than enters the moving bed regenerator. It is regenerated to produce sulfur dioxide and sodium sulfide by a counter-current stream of reducing gas. The reductant gas stream is a mixture of hydrogen and carbon monoxide. The sodium on the sorbent is present as the sulfide as it flows from the regenerator into a steam treatment vessel (Unit #10A). The steam reacts with the sodium sulfide to produce hydrogen sulfide.

The sorbent from the steam treatment reactor is lifted into a fluidized bed cooler where it meets a counter-current flow of cold air. From the cooler, it is returned to the adsorber by a pneumatic or bucket conveyer.

The cold air used in the fluid bed cooler is subsequently used as combustion air in the coal-fired air heater (Unit #14). In this unit, air is heated several hundred degrees above the regeneration temperature so that it can be used as the heating medium in the fluid bed heater (Unit #9). As it leaves the fluid bed heater, this gas, containing all the NO_x removed from the flue gas and a small amount of SO_2, is mixed with combustion air for the coal-fired boiler. Feeding this gas stream into the boiler combustion air provides for a method of recovering heat from the NOXSO system by raising the boiler combustion air temperature. It also provides a unique method for disposing of the NO_x collected. The normal operating temperature in a coal-fired boiler is between 1200 and 1300°C. At these temperatures, the incoming NO will decompose rapidly to N_2 and O_2. By the time the flue gas leaves the boiler, the NO concentration is close to its thermal equilibrium value of 510 ppm at 1200°C and 820 ppm at 1300°C, assuming the concentrations of N_2 and O_2 in the flue gas are 79% and 4%, respectively.

PROCESS ECONOMICS

During the past year, the Electric Power Research Institute (EPRI) initiated a design and economic study of the NOXSO process. NOXSO provided design information to the EPRI contractor. The design specifications provided by NOXSO for the major equipment items in a 125 MW_e NOXSO process module are presented in Table 4.

A draft of the EPRI cost study has been completed and was reviewed by NOXSO. In addition, NOXSO has obtained a design and cost review from two other independent organizations in the pollution control industry. These three independent studies form the basis of the cost data presented in this paper. A detailed description of this cost analysis is contained in a report recently submitted to the Department of Energy [NOXSO (1985)].

The NOXSO System capital and operating costs are summarized in Tables 5, 6, and 7. The costs pertain to two 500 MW_e flue gas treatment systems. One 125 MW_e process module encompasses an area of 6400 square feet and a height of 80 feet. The specifications for the coal-fired boilers used to develop these costs are for the high sulfur coal case contained in a recent EPRI report [EPRI (1983)].

For comparison, the NOXSO system costs are shown with the costs of comparable flue gas treatment systems in Table 8. All costs in Table 8 were

TABLE 4. Design Specifications of Major Equipment Items (125 MWe Module)

FLUIDIZED BED ADSORBER	
Cross sectional area	1990 ft^2
Settled bed height	1.84 ft
Bed temperature	120°C
Total pressure drop	22 in. H_2
Superficial gas velocity	2.95 ft/sec
Sorbent residence time	45.15 min.
Sorbent circulation rate	2.07×10^5 lb/hr
FLUIDIZED BED SORBENT HEATER	
Number of stages	3
Cross sectional area	765 ft^2
Pressure drop across reactor	41 in. H_2O
Inlet sorbent temperature	120°C
Outlet sorbent temperature	600°C
Inlet air temperature	694°C
Outlet air temperature	323°C
Amount of fluidization air required	1.16×10^3 mole/sec
MOVING BED REGENERATOR	
ID of reactor	15 ft
Height of reactor	8.6 ft
Inlet sorbent temperature	600°C
Pressure drop across reactor	45 in. of H_2
Requirement of reformed natural gas	2.39×10^3 SCFM
Conversion of Na_2SO_4 90%	
Sorbent residence time	18.7 min
Sulfur production rate	1.27×10^4 lb/hr
STEAM TREATMENT VESSEL	
ID of reactor	14 ft
Height of reactor	5 ft
Pressure drop across reactor	26 in. H_2O
Superficial velocity of steam	0.92 ft/sec
Steam requirement	8800 lb/hr
FLUIDIZED BED SORBENT COOLER	
Number of stages	3
Cross sectional area	503 ft^2
Pressure drop across reactor	41 in. H_2O
Inlet sorbent temperature	968°F
Outlet sorbent temperature	248°F
Inlet air temperature	108°F
Outlet air temperature	664°F
Amount of fluidization air required	1.16×10^3 mole/sec

TABLE 5. NOXSO System Total Capital Requirement*

Area	Description	$/kW
20	SO_2 Removal System	5.8
40	Regeneration System	37.0
30	Flue Gas Handling System	25.0
10	Reagent Feed System	1.4
60	Waste Handling System	8.3
50	By-Product System	38.0
80	Particulate Removal System	39.0
70	General	0.3
	Total Process Capital	155.0
General Facilities		15.5
Engineering and Home Office Fees		19.4
Project Contingency		32.5
Process Contingency		16.9
	Total Plant Cost	239.3
Allowance of Funds During Construction (AFDC)		8.9
	Total Plant Investment	248.0
Royalty Allowance		0.8
Preproduction Costs		10.2
Inventory Capital		11.7
Initial Catalyst and Chemicals		0.2
	Total Capital Requirement	271.0
	Less Capital for Particulate Control System	71.0
	Total Capital Requirement for FGD System	200.0

*December 1982 dollars, two 500 MW units.

TABLE 6. NOXSO System Operating Costs*

	1st Year $/kW-yr*	Levelized mills/kWh
Fixed Operating Costs		
Operating Labor	1.4	0.6
Maintenance Labor	2.4	1.0
Maintenance Material	3.6	1.4
Administration and Support Labor	1.2	0.5
Total Fixed Operating Costs	8.6	3.5
Variable Operating Costs		
Treated Alumina Sorbent	15.5	6.3
Coal	9.2	3.7
Liquid Oxygen	16.5	6.7
Cooling Water	0.03	0.01
Low Pressure Steam	0.2	0.08
Power	9.4	4.0
Dry Solids (unlined)	1.64	0.7
Catalyst - Acid Plant	0.02	0.01
Total Variable Operating Costs	52.5	21.5
Credits for By-Products	-16.2	-6.6
Total Variable Operating Costs with By-Products Credit	36.3	14.9

*December 1982 dollars, two 500 MW_e units.

TABLE 7. NOXSO System Levelized Cost*

	Mills/kWh
Process Capital	4.2
General Facilities	0.4
Engineering and Home Office Fees	0.5
Project Contingency	0.9
Process Contingency	0.5
Total Plant Cost	6.5
Allowance for Funds During Construction (AFDC)	0.2
Total Plant Investment	6.7
Royalty Allowance	0.02
Preproduction Costs	0.28
Inventory Capital	0.32
Total Capital Requirement	7.3
Fixed Operating Cost	3.5
Variable Operating Cost	21.5
Total Levelized Cost	32.3
Less By-Product Credit	-6.6
Total Levelized Cost including By-Product Credit	25.7
Reference Case Levelized Cost	3.8
Levelized Cost Apportioned to FGD	21.9

*Two 500 MW units.

TABLE 8. Cost Comparison for Flue Gas Treatment Systems*

	Capital Cost $/KW	Operating Costs Mills/KW hr.
THROWAWAY PROCESSES (FGD)		
Conventional Limestone	175	18
Chiyoda Thoroughbred 121	140	14
DOWA	175	14
Forced Oxidation of Limestone	175	16
Saarberg - Holter	130	16
Limestone Dual Alkali	160	16
Lime Dual Alkali	150	17
Wet Limestone	165	20
REGENERABLE PROCESSES (FGD)		
Wellman - Lord	275	26
MGO	270	19
Sulf-X	295	20
Flakt-Boliden	390	29
Aqueous Carbonate	400	30
Conosox	430	45
REGENERABLE PROCESSES (SO_2/NO_x)		
Avco-Ebara	275	27
NOXSO	200	22

*The costs are presented in December 1982 dollars using a standard set of EPRI cost analysis techniques.

derived by EPRI using a standard set of cost analysis techniques. The comparison shows that the NOXSO process is only 5% more expensive than conventional (limestone FGD) and 25% less expensive than other simultaneous NO_x/SO_2 processes and other regenerable FGD processes. Although slightly more expensive than limestone FGD, the NOXSO process has the advantage of removing NO_x and eliminating the production of large quantities of solid waste.

REFERENCES

Bowen, J.H. and C.K. Cheng, "Regeneration of Sulfated Alkalized Alumina," Environmental Science & Technology, 8(8), 747, (1974).

Deo, A.V.; I.G. Dalla Lana and H.W. Habgood, "Infrared Studies of the Adsorption and Surface Reactions of Hydrogen Sulfide and Sulfur Dioxide on Some Aluminas and Zeolites," J. Catal. 21, 270, (1971).

Glass, R.W. and R.A. Ross, "Surface Studies of Heat-Treated - Aluminas in the Adsorption of Sulfur Dioxide Below Monolayer Coverage," Canad. J. Chem., 50, 2537, (1972).

Goebel, J.A. and F.S. Pettit, Met. Trans., Vol. 1, pp. 1943-1954, 1970.

Goebel, J.A. and F.S. Pettit, Met. Trans., Vol. 1, pp. 3421-3429, 1970.

Khulbe, K.C. and R.S. Mann, "ESR Studies of SO_2 and H_2S Adsorption on Alumina and Alumina-Supported Mo and Mo-Co," J. Catal., 51, 364-371, (1978).

NOXSO Corporation, "Evaluation of the NOXSO Combined NO_x/SO_2 Flue Gas Treatment Process," Prepared for U.S. Department of Energy, Pittsburgh Energy Technology Center, Pittsburgh, Pa., September 30, 1983.

NOXSO Corporation, "Evaluation of the NOXSO Combined NO_x/SO_2 Flue Gas Treatment Process," Prepared for U.S. Department of Energy, Pittsburgh Energy Technology Center, Pittsburgh, Pa., July 6, 1984.

NOXSO Corporation, "Evaluation of the NOXSO Combined NO_x/SO_2 Flue Gas Treatment Process," Prepared for U.S. Department of Energy, Pittsburgh Energy Technology Center, Pittsburgh, Pa., April 1985.

Odin, L.L. and P.E. Francoeur, "Preparing Alkalized Alumina," U.S. Bureau of Mines, RI 7294, (1969).

Peri, J.B., "The Effect of Fluoride on Surface "Acid" Sites on Alumina and Silica-Alumina," J. Phys. Chem., 72, 2917, (1968).

Pozdnyakov, D.V. and V.N. Filimonov, "Use of IR Spectroscopy to Investigate Chemisorption of Nitric Oxide and Nitrogen Dioxide on Metallic Oxides," Kinet. Katal. (English Translation), 14, 655, (1973).

Roev, L.M. and A.V. Alekseev, "Elementary Photo Processes in Molecules," Moscow-Leningrad, (1966).

Singhal, S.C. (ed.), "High Temperature Protective Coatings," TMS-AIME, Warrendale, Pa., (1982).

Stearns-Rogers Engineering Corporation, "Economic Evaluation of FGD Systems," Vols. I, II, III, EPRI CS-3342, (1983).

Turnham, B.D., "The Physical and Chemical Adsorption of Nitric Oxide on Alumina and Other Oxides," Ph.D. Dissertation, Stanford University, (1975).

Yeh, J.T., U.S. Department of Energy, private communication, April 1985.

APPENDIX: Attendees

9th Symposium on Flue Gas Desulfurization

06/04/85-06/07/85
Netherland Plaza

Cincinnati, OH

List of Attendees

David W. Ablin
Sales Engineer
FMC Corp. - AQC
1501 Woodfield Rd., Ste. 300
E. Schaumburg, IL 60195
312/843-1700

Chuck Adami
Plant Manager
West Texas Utilities Co.
P.O. Box 841
Abilene, TX 79604
915/672-3251 X230

Andre Affatato
President
APA, Inc.
7809 Laurel Avenue #13
Cincinnati, OH 45243
513/561-0687

Giuseppe J. Aldina
System Specialist
Thermo Electron
105 South Street
Hopkinton, MA 01748
617/435-5321

Per-Olof Alfredsson
Senior Vice President
Flakt AB
Box 81 001
S-104 81 Stockholm Sweden
08/714 40 00

Robert A. Allan
Manager Engineering
Joy Mfg. Canada Ltd.
235 King Street East
Kitchener, Ontario CANADA
519/579-3965

Harold R. Althen
General Sales Manager
Peabody Process Systems, Inc.
201 Merritt 7
Norwalk, CT 06852
203/846-1600

D. B. Anderson
Marketing Manager
Laque Center Corrosion Tech.
Park 80
West Plaza Two, NJ 07662
201/368-4809

Jumpei Ando
Professor
Chuo University
1-13-27 Kasuga
Bunkyo-Ku Tokyo 112
JAPAN

Roger L. Antonie
Utility Sales Reg. Supervisor
Warman International, Inc.
2701 S. Stoughton Road
Madison, WI 53716
608/221-2261

Chuck Armstrong
President
Poly Manufacturing
Hwy. 59
North Tenaha, TX 75974
409/248-2903

Paul A. Arsenault
Sr. Project Engineer
Public Service Co. of NH
1000 Elm Street
Manchester, NH 03105
603/669-4000 X2570

Sachdev Ashok
Research Engineer
Atomic Energy of Canada Ltd.
Pinawa MB
ROE IL0 Canada
204/753-2311 X 2438

Franklin A. Ayer
Consultant
Research Triangle Inst.
P.O. Box 12194
Research Triangle Park, NC 27709
919/541-6260

Ronald J. Bacskai
President
Conversion Systems, Inc.
115 Gibraltar Road
Horsham, PA 29044
215/441-5900

Michael P. Bahor
Engineering Manager
GAI Consultants, Inc.
570 Beatty Rd.
Monroeville, PA 15146
412/856-6400

Robert J. Bailey
product Manager
Price Brothers Co.
P.O. Box 825
Dayton, OH 45401
513/226-8706

Even Bakke
V.P. Research & Devel.
Peabody Process Systems, Inc.
201 Merritt 7
Norwalk, CT 06852
203/846-1600

Milton V. Balch
Project Planning Engineering
Southern Company Services
P.O. Box 2625
Birmingham, AL 35202
205/877-7297

Peter Baldwin
V.P. & Gen. Mgr. Single Stage Prod.
Ingersoll-Rand Company
5510 77 Center Drive
Charlotte, NC 28224
704/527-0500

H. A. Bambrough
Head, Major Combustion
Canadian Federal Govnt.
351 Blvd. St. Joseph
Hull, Quebec, Canada
819/997-1220

Charles R. Barden
Consulting Engineer
Hydro-Sonic
8611 Alverstone Way
Austin, TX 78759
512/343-2381

R. T. Beall
V.P. & Sales Manager
Gifford-Hill & Co., Inc.
300 E. John W. Carpenter Frwy.
Irving, TX 75062
214/258-7321

Bruce Beckwith
Project Manager
Pfaudler Company
1000 West Avenue
Rochester, NY 14692
716/464-4898

Bob Bedaw
Lab Technician
Bete Fog Nozzle, Inc.
324 Wells Street
Greenfield, MA 01301
413/772-0846

R. P. Bell
President
Monserco
6620 Kitmat Road
Mississauga, Ontario
416/821-1033

S. Bengtsson
n/a
Flakt Industrial, AB, Sweden
P.O. Box 87
Knoxville, TN 37901
615/693-7550

Lewis B. Benson
Technical Serv. Engr.
Dravo Lime
3600 Neville Road
Pittsburgh, PA 15225
412/777-5572

Al Berst
Manager Engineering
Zurn Industries, Inc.
245 Center Street North
Birmingham, AL 35226
205/252-2181

Greg Betenson
Engineer
Utah Power & Light Co.
P.O. Box 899
Salt Lake City, UT 84110
801/535-4915

John Bevilacquia
Product Specialist
Sohio Engineered Mtls. Co.
P.O. Box 1054
Niagara Falls, NY 14302
N/A

Ananda P. Bhat
Generation Process Eng'g Mgr.
Pennsylvania Electric Co.
1001 Broad Street
Johnstown, PA 15907
814/533-8560

Vinay K. Bhatia
Engineer
F. L. Smidth & Co. A/S
22 Hoffdingsvej 2500 Valby
Denmark
00945 1 301166

Frederick J. Bickerton, Jr.
Engineer & Safety Coordinator
Duquesne Light Co.
Phillips Power Station
So. Heights, PA 15081
412/457-7510

Edward L. Biedell
Manager FGD Applications
MikroPul Corporation
10 Chatham Road
Summit, NJ 07901
201/273-6360

Gregory Bielawski
FGD Product Mgr.
Babcock & Wilcox
P.O. Box 351
Barberton, OH 44281
216/860-1591

Robert J. Biolchini
Process Engineer
FMC Corp. - AQC
1501 Woodfield Rd., Ste. 300
E. Schaumburg, IL 60195
312/843-1700

John R. Bitler
Director, Commercial Dev.
Northern Natural Resources
2223 Dodge St.
Omaha, NE 68102
402/691-2290

Steven A. Bjorklun
n/a
Burns & McDonnell
P.O. Box 173
Kansas City, MO 64141
816/333-4375

Gary Blythe
Sr. Chemical Engineer
Radian Corporation
P.O. Box 9948
Austin, TX 78766
512/454-4797

Steven L. Boes
Bradleygineer
Central & South West Services
P.O. Box 1600
Vernon, TX 76384
817/886-1651

David C. Borsare
Chief Engineer
Combustion Engineering, Inc.
31 Inverness Center Parkway
Birmingham, AL 35243
205/967-9100

William H. Boss
Senior Research Engineer
Univ. of Tenn. Space Inst.
n/a
Tullahoma, TN 37388
615/455-0631 X363

Edmond J. Boucher
Ex. Vice President
ABCO Plastics, Inc.
136 Simsbury Road
Avon, CT 06001
203/678-0720

Willard L. Boward
Manager, Process Engineering
FMC Corp. - AQC
1501 Woodfield Rd., Ste. 300
E. Schaumburg, IL 60195
312/843-1700

Ralph Boyd
Salesman
Black River Lime Co.
P.O. Box 1
Butler, KY 41006
606/472-7721

Wallace H. Bradley
V.P., Engineering
Austell Box Board Corp.
3100 Washington St.
Austell, GA 30001
404/948-3100

Jon A. Brander
Application Engineer
General Electric
1 River Road, Bldg. 2
Schenectady, NY 12345
518/385-9894

Richard Brelowski
Technical Manager
Steetley Resources Inc.
128 E. Main
Woodville, OH 43469
419/849-2321

Theodore G. Brna
Program Manager
U.S. EPA
IERL, MD-61
Research Triangle Park, NC 27711
919/541-2683

Gordon Brown
Marketing Manager
Morrison Molded Fiber Glass Co.
400 Commonwealth
Bristol, VA 24201
703/669-1181

Robert M. Brown
Generation Sales Engineer
General Electric
P.O. Box 68543
Indianapolis, IN 46268
317/298-5209

Van D. Brugghen
Chemical Engineer
Kema
Utrechtseweg 310
6812 Ar Arnhem, NETHERLANDS
085-562568

Phillip Brundrett
Mgr. Market Development
W. R. Grace-Davison Chem. Div.
110 E. Baltimore St.
Baltimore, MD 21203
301/659-9192

William F. Buerger
Power Ind. Co-or.
Betz Labs, Inc.
260 Northland, Ste 206
Cincinnati, OH 45246
513/772-0060

David Burford
Sr. Research Engineer
Southern Company Services
P.O. Box 2625
Birmingham, AL 35202
205/870-6329

Jack M. Burke
n/a
Radian Corporation
P.O. Box 9948
Austin, TX 78766
512/454-4797

Richard Burmester
Vice President
The Clarkson Company
3430 West Bayshore Road
Palo Alto, CA 94303
415/494-1010

David Burrichter
Senior Project Engineer
Armstrong World Inst.
P.O. Box 3511
Lancaster, PA 17604
717/396-5854

Vann Bush
Head, Applied Physics Section
Southern Research Institute
P.O. Box 55305
Birmingham, AL 35255
205/323-6592

C. Butler
Senior Environment Engr.
Corning Glass
HPME-1
Corning, NY 14831
607/974-6683

Barry Buttler
Staff Engineer
Krebs Engineers
1205 Chrysler Drive
Menlo Park, CA 94025
415/325-0751

Ronald M. Camp
Cost Analyst
Southern Co. Services
800 Shades Creek Pkwy.
Birmingham, AL 35202
205/870-6442

Ivor E. Campbell
N/A
Ivor E. Campbell & Associates
P.O. Box 153
New Albany, OH 43054
614/855-2183

Bernard S. Camponeschi
Industry Manager
FMC Corp. - AQC
1501 Woodfield Rd., Ste. 300
E. Schaumburg, IL 60195
312/843-1700

Robert Candelaria
Lead Environmental Engr.
Salt River Project
P.O. Box W
Page, AZ 86040
602/645-8811

Joseph Cannon
Partner
Pillsbury, Madison and Sutro
1667 K Street NW, Ste. 1100
Washington, DC 20006
202/887-0300

W. R. Cares
Sr. Research Chemist
M. W. Kellogg
16200 Park Row
Houston, TX 77084
713/492-2500

Jim Carneal
Mgr.-Property & Environ. Affairs
Louisville Gas & Electric Co.
311 W. Chestnut Street
Louisville, KY 40232
502/566-4612

John K. Carpenter
Supervising Engineer
Union Electric Co.
P.O. Box 149
St. Louis, MO 63166
314/554-2911

Teri Carter
Process Engineer
Conoco, Inc.
P.O. Box 1267
Ponca City, OK 74603
405/767-3131

Lewis A. Cathey
Supervising Engineer
Houston Lighting & Power
12301 Kurland Drive
Houston, TX 77034
713/481-7897

Frank S. Chalmers
Consultant
Chalmers & Company
2022 Columbia Rd., NW #401
Washington, DC 20009
202/331-1248

John C.S. Chang
Senior Chemical Engineer
Acurex Corporation
P.O. Box 13109
Research Triangle Park, NC 27709
N/A

Edward L. Chapman
Regional Manager
FMC Corp. - AQC
1501 Woodfield Rd., Ste. 300
E. Schaumburg, IL 60195
312/843-1700

Art Chapple
Southeastern Sales Mgr.
LTV/Fibercast
P.O. Box 467877
Atlanta, GA 30346
404/452-8249

Yang-Jen Chen
Sr. Research Engineer
Joy Manufacturing Co.
4565 Colo Blvd.
Los Angeles, CA 90039
818/240-2300

Gregory H. Cheng
Technical Director
The Ducon Company
147 E. 2nd Street
Mineola, NY 11501
516/741-6100

Sumas Chitnis
Manager of Engineering
General Electric Env. Services
200 N. 7th St.
Lebanon, PA 17042
717/397-8812

Ove B. Christiansen
n/a
Niro Atomizer
9165 Rumsey Road
Columbia, MD 21045
301/997-8700

Roger Christman
n/a
Christman & Associates
11708 Bowman Green Drive
Reston, VA 22090
703/435-3219

Paul Chu
n/a
University of Texas
Dept. of Chemical Engineering
Austin, TX 78712-1165
512/471-7230

Edwarad Cichanowicz
Project Manager
Electric Power Research Inst.
P.O. Box 10412
Palo Alto, CA 94303
415/855-2374

David R. Clark
National Manager
Allied Corp.
8825 Weller Road
Cincinnati, OH 45249
513/489-5340

Curtis W. Clarkson
President
The Clarkson Co.
3430 W. Bayshore
Palo Alto, CA 94303
415/494-1010

Donald D. Clasen
Marketing Manager
Chiyoda International Corp.
1300 Park Place Bldg.
Seattle, WA 98101
206/624-9350

David E. Clay
Export Marketing Manager
Kraftanlagen Aktiengesellschaft
IM Breitspiel 7
6900 Heidelberg 1, GERMANY
06221-394335

Mary B. Cleary
Chemist
High Technology Coatings
3838 D. W.O.W.
Corpub Christi, TX 78415
512/857-0228

John R. Cline
Proposal Engineer
Babcock & Wilcox
20 S. Van Buren
Barberton, OH 44203
216/860-2019

Edwin H. Cloth
Supv., Materials Engineering
Stone & Webster Engineering Corp.
P.O. Box 5200
Cherry Hill, NJ 08034
N/A

Robert Cmiel
Engineering Supervisor
San Miguel Electric
P.O. Box 280
Jourdanton, TX 78026
512/784-3411

Alyson Cobb
Plant Engineer
Jacksonville Electric Authority
11201 New Berlin Road
Jacksonville, FL 32226
904/766-5494

David Colley
n/a
Radian Corp.
P.O. Box 9948
Austin, TX 78766
512/454-4797

Thomas F. Conlan
Secretary
Kissick Corp.
108 Benson East
Jenkintown, PA 19046
215/885-6650

Jack G. Conner
Marketing Manager
Battelle Memorial Inst.
505 King Avenue
Columbus, OH 43201
614/424-4183

James Cooper
Senior Engineer
James Howden & Co. Ltd.
195 Scotland Street
Glasgow, Scotland G58PJ
041/429-2131

Edwarad Cichanowicz
Project Manager
Electric Power Research Inst.
P.O. Box 10412
Palo Alto, CA 94303
415/855-2374

David R. Clark
National Manager
Allied Corp.
8825 Weller Road
Cincinnati, OH 45249
513/489-5340

Curtis W. Clarkson
President
The Clarkson Co.
3430 W. Bayshore
Palo Alto, CA 94303
415/494-1010

Donald D. Clasen
Marketing Manager
Chiyoda International Corp.
1300 Park Place Bldg.
Seattle, WA 98101
206/624-9350

David E. Clay
Export Marketing Manager
Kraftanlagen Aktiengesellschaft
IM Breitspiel 7
6900 Heidelberg 1, GERMANY
06221-394335

Mary B. Cleary
Chemist
High Technology Coatings
3838 D. W.O.W.
Corpub Christi, TX 78415
512/857-0228

John R. Cline
Proposal Engineer
Babcock & Wilcox
20 S. Van Buren
Barberton, OH 44203
216/860-2019

Edwin H. Cloth
Supv., Materials Engineering
Stone & Webster Engineering Corp.
P.O. Box 5200
Cherry Hill, NJ 08034
N/A

Robert Cmiel
Engineering Supervisor
San Miguel Electric
P.O. Box 280
Jourdanton, TX 78026
512/784-3411

Alyson Cobb
Plant Engineer
Jacksonville Electric Authority
11201 New Berlin Road
Jacksonville, FL 32226
904/766-5494

David Colley
n/a
Radian Corp.
P.O. Box 9948
Austin, TX 78766
512/454-4797

Thomas F. Conlan
Secretary
Kissick Corp.
108 Benson East
Jenkintown, PA 19046
215/885-6650

Jack G. Conner
Marketing Manager
Battelle Memorial Inst.
505 King Avenue
Columbus, OH 43201
614/424-4183

James Cooper
Senior Engineer
James Howden & Co. Ltd.
195 Scotland Street
Glasgow, Scotland G58PJ
041/429-2131

David R. Cosper
Research Associate
Nalco Chemical Co.
1601 W. Diehl Road
Naperville, IL 60566
312/961-9500

Eric Costello
Principle Engineer
Tampa Electric Co.
P.O. Box 111
Tampa, FL 33601
813/677-9141

John W. Coulston
Environmental Control Specialist
Spraying Systems Co.
North Ave. at Schmale Rd.
Wheaton, IL 60188
312/665-5000

Bob Cowen
President
Martek, Inc.
85 Research Road
Hingham, MA 02043
617/749-6992

Ellis B. Cowling
Associate Dean for Research
North Carolina State Univ.
P.O. Box 8001
Raleigh, NC 27695
919/737-2883

Thomas Cox
Operations Supervisor
Muscatine Power & Water
3205 Cedar Street
Muscatine, IA 52761
319/263-2631

A. B. Craig
Deputy Director
U.S. EPA
MD-60
Research Triangle Park, NC 27711
919/541-2821

Ken Craig
Asst. Supt. S.W.P.P.
City Utilities
301 E. Central
Springfield, MO 65802
417/882-5075

Thomas Crandall
Technical Sales
Stebbins Engineering
363 Eastern Blvd.
Watertown, NY 13601
315/782-3000 X224

George E. Cranston
President
Peabody Process Systems, Inc.
201 Merritt 7
Norwalk, CT 06852
203/846-1600

William R. Cress
Manager, Engineering Studies
Allegheny Power Service Corp.
800 Cabin Hill Drive
Greensburg, PA 15601
412/838-6721

Thomas Cummins
Operating Engineer
CG&E
4th & Main
Cincinnati, OH 45201
513/553-7517

Jim Cunagin
Technical Specialist
Betz Labs
6305 Two Springs Lane
Louisville, KY 40207
513/772-0060

John D. Cunic
Senior Staff Engineer
Exxon Research & Engineering Co.
P.O. Box 101
Florham Park, NJ 07932
201/765-3986

Albert R. Cunningham
Senior Engineer
Philadelphia Electric Co.
2301 Market Street
Philadelphia, PA 19101
215/841-5046

Robert S. Dahlin
Senior Chemical Engineer
Southern Research Institute
2000 Ninth Avenue S
Birmingham, AL 35255
205/323-6592

Stuart M. Dalton
Program Manager
Electric Power Research Inst.
P.O. Box 10412
Palo Alto, CA 94303
415/855-2467

Ashok Damle
Research Engineer
Research Triangle Inst.
P.O. Box 12194
Research Triangle Park, NC 27709
919/541-5829

Ray Danz
Operations Superintendent
Muscatine Power & Water
3205 Cedar Street
Muscatine, IA 52761
319/263-2631

Daniel M. Dasho
Public Service Engineer 4
Public Service Comm. of WI
P.O. Box 7854
Madison, WI 53707
608/266-0910

Richard L. Davis
Man. Engr. & Tech. Dir.
Koppers Co., Inc.
2701 Koppers Bldg.
Pittsburgh, PA 15243
412/227-2610

Wayne T. Davis
Professor
University of Tennessee
Room 73A Perkins Hall
Knoxville, TN 37996-2010
615/974-7728

John C. De Ruyter
Senior Engineer
E. I. du Pont
Louviers 33E113
Wilmington, DE 19898
302/366-6442

Larry Decker
Senior Staff Engineer
Dynatech RID
99 Erie Street
Cambridge, MA 01701
617/868-8050

Bill Delis
Supt. of Operations
Cin. Gas & Electric
East Bend Station
Union, KY 41091
606/586-5600

G. Della Faille
n/a
Solvay & Cie, S.A.
609 5th Avenue, Room 701
New York, NY 10017
212/838-8563

Dean Delleney
Program Manager
Radian Corp.
8501 Mo-Pac Blvd.
Austin, TX 78766
512/454-4797

Herbert Dempsey
Project Manager
Acurex Corp.
P.O. Box 13109
Research Triangle Park, NC 2709
919/544-4535

Chuck Dene
Project Manager
Electric Power Research Inst.
P.O. Box 10412
Palo Alto, CA 94303
415/855-2425

William DePriest
Environmental Systems
Babcock & Wilcox
Barberton, OH 44203
N/A
216/860-2677

Prakash H. Dhargalkar
FGD Product/Applications Mgr.
Research-Cottrell
P.O. Box 1500
Somerville, NJ 08876
201/685-4295

N. N. Dharmarajan
Sr. Engineer Emission Control
Central & Southwest Services, Inc.
P.O. Box 660164
Dallas, TX 75266-0164
214/754-1373

Jim Dickerman
Senior Program Manager
Radian Corporation
P.O. Box 13000
Research Triangle Park, NC 27709
919/541-9100

Douglas Dieterich
Project Manager
Procter & Gamble Paper
P.O. Box 8020
Green Bay, WI 54304
414/433-2707

Richard A. Diffenbach
n/a
Dept. of Energy
P.O. Box 10940
Pittsburgh, PA 15236
412/675-6090

Eldon Dille
n/a
Burns & McDonnell
P.O. Box 173
Kansas City, MO 64141
816/333-4375

Daniel A. Dillman
Chemical Services Sup.
Duquesne Light Co.
2101 Beaver Ave.
Pittsburgh, PA 15233
412/393-4896

John Dilzell
Marketing Manager
Betz Laboratory
4636 Somerton Road
Trevose, PA 19047
215/355-3300

Richard L. Dirr
Principal Mechanical Eng.
Cincinnati Gas & Electric Co.
P.O. Box 960
Cincinnati, OH 45201
513/632-3462

Richard D. Doctor
Chemical Engineer
Argonne National Laboratory
9700 S. Cass
Argonne, IL 60439
312/972-5913

Jim Donnelly
Mgr. APC R&D Process
A/S Niro Atomizer
Gladsaxevej 305
DK2860 Soeborg, Denmark
(45) 1-691011

Thomas E. Dowdy
Sr. Chemical Engineer
U. T. Space Institute
n/a
Tullahoma, TN 37388
615/455-0631 X410

William Downs
Research Specialist
Babcock & Wilcox Company
1562 Beeson Street
Alliance, OH 44601
216/821-9110 X359

Antonio J. DoVale, Jr.
Chief Process Engr., FGD Systems
Wheelabrator Air Pollution Control
600 Grant Street
Pittsburgh, PA 15219
412/288-7378

Jerry Dubiansky
District Manager
MPSI
200 Fleet Street
Pittsburgh, PA 15220
412/922-5252

Ronald G. Duffy
Sales manager-FGD
Peabody Process Systems, Inc.
201 Merritt 7
Norwalk, CT 06852
203/846-1600

Eddie D. Duke
Vice President
Poly Meg., Inc.
P.O. Box 158
Tenaha, TX 75974
409/248-2903

Fitzroy A. Dyer
Project Manager
Church & Dwight Co., Inc.
20 Kingsbridge Road
Piscataway, NJ 08854
201/-885-1220

Mark A. Dyer
Environmental Engineer
Central Ill. Public Service Co.
607 East Adams St.
Springfield, IL 62701
217/523-2600 X847

Kenneth W. Earle
Shift Engineer
Tennesses Valley Authority
P.O. Box 800
Drakesboro, KY 42337
502/476-2250

James Eddinger
Environmental Engineer
U.S. EPA
Research Triangle Park,
NC 27712
919/541-5595

Eleanor Edwards
Proposal manager
Babcock & Wilcox
20 S. Van Buren Ave.
Barberton, OH 44203
216/860-2380

Randy Edwards
East Central Dist. Sales Mgr.
ABCO Plastics, Inc.
136 Simsbury Road
Avon, CT 06001
203/678-0720

Richard T. Egan
Product Engineering Manager
The Munters Corp.
P.O. Box 6428
Fort Myers, FL 33907
813/936-1555 X131

Brooke Eldredge
Conference & Travel Coordinator
Electric Power Research Inst.
P.O. Box 10412
Palo Alto, CA 94303
415/855-7919

Bill Ellison
Director
Ellison Consultants
4966 Tall Oaks Drive
Monrovia, MD 21770
301/865-5302

Thomas E. Emmel
Senior Chemical Engineer
Radian Corp.
P.O. Box 13000
Research Triangle Park, NC 27709
919/541-9100

Bruce Epperson
Mechanical Engineer
Tennessee Valley Authority
W3A41, 400 Summit Hill Dr.
Knoxville, TN 37902
615/632-3390

Olav Erga
Professor
Norw. Inst. of Techn.
Sem Saelandsvei 4
Trondheim NORWAY
07/594134

Michael Esche
Dip.-Ing, Dipl.-Wirtsch. Ing.
Saarberg-Holter Umwelttechnik GmbH
Hafenstrasse 6, 6600 Saarbrucken
West Germany
0681/32104-06

Arnold L. Evans
Manager New Applications
Allied Corp.
P.O. Box 1559
Hopewell, VA 23860
804/541-5660

Earl Evans
Project Manager
U.S. Dept. of Energy
P.O. Box 10940
Pittsburgh, PA 15236
412/675-6237

William W. Farmer
Asst. Section Supv. Engineer
Georgia Power Company
P.O. Box 4545
Atlanta, GA 30306
404/526-7194

John H. Fassnacht
Sr. Territory Manager
DuPont
N-9451
Wilmington, DE 19898
302/774-7863

Daniel Fauth
Research Chemist
U.S. Dept. of Energy Petc.
P.O. Box 10940
Pittsburgh, PA 15236
412/675-6093

Howard Feibus
Dir., Office of Coal Utilization
U.S. DOE, Coal Utility Systems
MS F323
Washington, DC 20545
301/353-4348

Bryan W. Ferguson
Chemical Engineer
Texas Utilities Generating Co.
400 North Olive St. LB-81
Dallas, TX 75201
214/979-8721

William B. Ferguson, Jr.
Dry Scrubbing Proposal Mgr.
Combustion Engineering
31 Inverness Ctr. Pkwy.
Birmingham, AL 35243
205/967-9100

Frank Ferraro
Environmental Program Mgr.
Babcock & Wilcox
20 S. Van Buren
Barberton, OH 44203
216/860-6781

Greg C. Ficke
Asst. Mgr.-Licensing & Env. Affairs
Cincinnati Gas & Electric Co.
P.O. Box 960
Cincinnati, OH 45201
513/632-2672

Carl E. Fink
Sr. Process Engineer
Conoco Coal Research
4000 Brownsville Rd.
Library, PA 15129
412/854-6581

M. T. Fisher
Pulp Engineering Group Leader
Procter & Gamble
P.O. Box 8010
Green Bay, WI 54308-8020
414/433-2731

Harold R. Fletcher
Air Pollution Control Equip. Engr.
Detroit Edison Company
2000 Second Avenue
Detroit, MI 48226
313/897-13151

Tim Flora
Plant Supervisor
Ohio Edison Edgewater Plant
200 Oberlin Ave.
Lorain, OH 44052
216/244-1991 X218

R. D. Forbus
Project Manager
Central & South West Serv.
P.o. Box 660164
Dallas, TX 75266-0164
214/754-1365

George C. Ford
President
George C. Ford & Asso.
607 Rosedale Road
Princeton, NJ 08540
609/924-0601

Fredrick W. Forsythe
Project manager
M. W. Kellogg
3 Greenway Plaza
Houston, TX 77046
713/960-4135

R. C. Forsythe
Mgr. Utility Sales & Service
Dravo Lime Company
3600 Neville Road
Pittsburgh, PA 15225
412/777-5572

Nicholas B. Franco
Director of Marketing
Bethlehem Mines Corp.
1791 Martin Tower
Bethlehem, PA 18016
215/694-3019

Peter E. Frankenburg
Research Associate
E. I. DuPont De Nemours & Co.
701 Bldg., CR
Wilmington, DE 19888
302/999-4858

Bill Frazier
Staff Engineer
Virginia Power
P.O. Box 26666
Richmond, VA 23261
804/775-5285

John G. Frederiksen
Business Development Mgr.
Monsanto
800 N. Lindbergh
St. Louis, MO 63167
314/694-8373

Douglas N. Freeman
Process Engineer
MEI Systems, Inc.
3121 W. Spencer
Appleton, WI 54914
414/725-7752

Michael Freholm
Sales Engineer
Borg Warner Industrial Prod.
P.O. Box 210
Ashland, KY 41101
606/319-0115

Dan Froelich
Manager, Air Quality Design
Burns & McDonnell Engineering
P.O. Box 173
Kansas City, MO 64141
816/333-4375

Robert Froment
Process Engineer
NEV
Rue de College
Marq en Baroeul FRANCE
(20) 72-1700

Cynthia Gage
n/a
University of Texas
Dept. of Chemical Engineering
Austin, TX 78712-1165
512/471-7230

Joseph Gaines
Senior Staff Engineer
Rust International Corp.
P.O. Box 101
Birmingham, AL 35201
205/930-2054

Edward E. Galloway
Director Environmental Affairs
Cincinnati Gas & Electric Co.
P.O. Box 960
Cincinnati, OH 45201
513/632-2268

John Garlanger
n/a
Ardaman and Associates, Inc.
P.O. Box 13003
Orlando, FL 32809
305/855-3860

Robert Gaugh
Senior Staff Engr.
Armco, Inc.
703 Curtis Street
Middletown, OH 45053
513/425-2488

Donald Gaugherbaugh
Sales Engineer
Kennedy Van Saun
350 Railroad St.
Danville, PA 17821
717/275-3050

John Gaynor
Group Leader
USG Corporation
700 W. Hwy. 45
Libertyville, IL 60048
312/362-9797

Gary E. Geiger
Chemist
Betz Labs
Somerton Road
Trevose, PA 19047
215/355-3300

Arthur L. Genoble
Environmental Engineer
Daniel International
Daniel Bldg.
Greenville, SC 29602
803/281-4480

Carl R. Gerber
Director
U.S. EPA
n/a
Washington, DC 20460
N/A

John R. Ghiringhelli
Construction Division Director
East Kentucky Power Cooperative
P.O. Box 707
Winchester, KY 40391
606/744-4812

James B. Gibson, Jr.
Mechanical Engineer
Tennessee Valley Authority
W7D194, 400 West Summitt Hill Dr.
Knoxville, TN 37902
615/632-4626

Douglas Gifford
n/a
Bell Tech
1129 Bellwood Avenue
Bellwood, IL 60104
312/547-7840

Srikant Gir
Engineer
Kerr-McGee
P.O. Box 25861
Oklahoma City, OK 73125
405/641-8551

Trebbi Giuliano
Head of Dept. of Fossil Fuel
ENEL/Italy
405 Lexington Avenue
New York, NY 10017
212/682-8150

George P. Gladis
Manager, Technical Services
Carpenter Technology Corp.
Springfield Rd.
Union, NJ 07083
201/686-7230

John Glamser
n/a
Dresser Industries
601 Jefferson
Houston, TX
713/391-1114

Cynthia Gleason
n/a
University of Texas
Dept. of Chemical Engineering
Austin, TX 78712-1165
512/471-7230

Rob Glover
n/a
Electric Power Research Institute
P.O. Box 10412
Palo Alto, CA 94303
415/855-2000

David Goettel
Environmental Supervisor
Santee Cooper
P.O. Box 1275
Georgetown, SC 29442
803/546-4171 X41

Dean M. Golden
Project Manager, Solid By-Product
Electric Power Research Inst.
P.O. Box 10412
Palo Alto, CA 94303
415/855-2516

Eduardo Gonzales
Supv. Environmental Engr.
Public Service of Colo.
5900 E. 39th Avenue
Denver, CO 80209
303/329-1663

Richard W. Goodwin
President
ECA, Inc.
14 Ramapo Lane
Saddle River, NJ 07458
201/934-9866

Jack Goudy
Business Development Mgr.
General Electric Co.
200 High Tower Blvd.
Pittsburgh, PA 15205
412/566-4073

Rolf Graf
n/a
Lurgi Corporation
666 Kinderkamack Road
River Edge, NJ 07661
201/967-7700

Richard J. Grant
Manager, Environmental Affairs
Central Ill. Public Serv. Co.
607 East Adams St.
Springfield, IL 62701
217/525-5367

Greg Graves
Project Engineer
Burns & McDonnell Engineering
P.O. Box 173
Kansas City, MO 64141
816/333-4375

Joseph Greco
Sales Manager
The Taulman Sales Co.
P.O. Box 52383
Atlanta, GA 30355
404/261-2535

Clois L. Green
Environmental Manager
Alcoa
P.O. Box 472
Rockdale, TX 76567
512/446-8205

Jack Greene
Administrative Officer
U.S. EPA
MD-60
Research Triangle Park, NC 27711
919/541-2903

Gary W. Grubbs
Electrical Engineer
Santee Cooper
One Riverwood Drive
Moncks Corner, SC 29461
803/761-4038

Bill Guernsey
Development Engineer
TRW - Energy Division
One Space Park
Redondo Beach, CA 90278
213/517-7337

John B. Guernsey
Vice President, Technical
Eastern Stainless Steel Co.
P.O. Box 1975
Baltimore, MD 21203
301/522-6428

David G. Guetig
Engineer
Indianapolis Power & Light Co.
RR #1
Petersburg, IN 47567
812/354-8801

Albert Hackl
Prof. Dipl.-Ing. Dr.
Technical University Vienna
Getreidemarkt 9
1060 Vienna - Austria
(222) 5601 X4726

B. O. Haeberlin
n/a
Louisville Gas & Elec. Co.
311 W. Chestnut Street
Louisville, KY 40232
502/566-4612

Wayne F. Hall
V.P. Product Development
Dravo Corporation
One Oliver Plaza
Pittsburgh, PA 15222
412/566-3643

Doug Hammontree
Project Manager
Burns & McDonnell Engineering
P.O. Box 173
Kansas City, MO 64141
816/333-4375

Svend Hansen
Vice President
Niro Atomizer, Inc.
N/A
Columbia, MD 21012
301/997-8700

O. W. Hargrove
n/a
Radian Corporation
P.O. Box 9948
Austin, TX 78766
512/454-4797

Richard C. Harrington
Sr. Engineer
Cleveland Electric Illuminating
P.O. Box 5000
Cleveland, OH 44101
216/622-9800 X3258

Peter Harriott
Professor
Cornell University
101 Olin Hall
Ithaca, NY 14853
607/256-3525

Harry F. Hart
Global Manager
Dorr-Oliver
77 Havemeyer Lane
Stamford, CT 06904
203/358-3442

James Harvilla
Project Engineer
New York State Elec. & Gas
4500 Vestal Parkway
Binghamton, NY 13903
607/729-2551 X4202

Seyed N. Hashemi
Engineer
Potomac Electric Co.
1900 Pennsylvania Ave. N.W.
Washington, DC 20068
202/331-6280

N. Hashimot
n/a
Chiyoda Chemical Engr. & Const.
1300 Park Place Bldg.
Seattle, WA 98101
206/624-9350

John Haslbeck
Program Manager
Noxso Corporation
P.O. Box 586
Library, PA 15129
412/675-6293

Sheila W. Hedges
n/a
U.S. Dept. of Energy
P.O. Box 10940
Pittsburgh, PA 15236
412/675-4614

Mel Henderson
Sales Engineer
U.S.G. Industries, Inc.
739 Fruithurst Drive
Pittsburgh, PA 15228
412/343-6044

Larry J. Henson
Chemical Engineer
Tennessee Valley Authority
501 CEB
Muscle Shoals, AL 35661
205/386-2514

Robert A. Hentges
Senior Engineer
Procter & Gamble Co.
6105 Center Hill Road
Cincinnati, OH 45224
513/659-5787

Albert A. Herring
Southeast Dist. Sales Mgr.
ABCO Plastics, Inc.
136 Simsbury Road
Avon, CT 06001
203/678-0720

H. F. Hess
Operation Manager
FMC Corp. - AQC
1501 Woodfield Road, Suite 300
East Schaumburg, IL 60195
312/843-1700

R. A. Hewitt
Supervisory Engineer
Texas Utilities Generating Co.
400 N. Olive
Dallas, TX 75201
214/979-8706

Wayne Hickok
Senior Production Engineer
Cooperative Power Association
14615 Lone Oak Road
Eden Prairie, MN 55344
612/937-8599

Robert G. Hilton
Director, Sales Development
Conversion Systems, Inc.
115 Gibraltar Road
Horsham, PA 19044
215/441-5900

Mike Hirst
Environmental Specialist
Tippett & Gee, Inc.
502 N. Willis
Abilene, TX 79603
915/673-8291

Art Hoekstra
Chemical Engineer
consultant
906 N. Woodlyn
Cincinnati, OH 45230
513/231-5558

K. B. Holland
Senior Engineer
James Howden Aust. P/L
97-103 Pacific Highway
North Sydney NSW 2060
02-9294566

O. L. Holland
Operations Mgr.
Hydrosonics Systems
P.O. Box 97
Lone Star, TX 75668
214/656-6317

Gerald A. Hollinden
Branch Chief
Tennessee Valley Authority
1130 Chestnut Street Towers II
Chattanooga, TN 37401
615/751-3584

David B. Holstein
Engineer
Texas Utilities Generating Co.
400 N. Olive St.
Dallas, TX 75201
214/979-8772

Charles C. Hong
Senior Research Engineer
Columbia Gas System
1600 Dublin Road
Columbis, OH 43215
614/481-1454

Richard G. Hooper
Project Manager
Electric Power Research Inst.
P.O. Box 10577
Denver, CO 80210
303/936-2981

Kevin Hoover
Service Supervisor
General Electric Co.
1150 Century Cir. No.
Springdale, OH 45246
513/671-2770

Bill Horton
Chemical Engineer
Lower Colo. River Authority
P.O. Box 220
Austin, TX 78767
512/473-3544

Richard Houlihan
Process Engineer
E.R.C.B.
640 5th Avenue
Calcary, Alberta, CANADA
403/297-3235

Louis S. Hovis
Chemical Engineer
U.S. EPA
Gas Cleaning Division
Research Triangle Park, NC 27111
919/541-3374

Michael Hudachko
Sales Engineer
FMC Corp. - AQC
1501 Woodfield Rd., Ste. 300
E. Schaumburg, IL 60195
312/843-1700

Wayne Huebner
Technical Director
Wallace & Tiernan
25 Main Street
Bellville, NJ 07109
201/759-8000 X438

Francis R. Hull
Consulting Engineer
Francis R. Hull, P.E.
23-03 45th Road
Long Island, NY 11101
718/937-1156

Thomas B. Hurst
Manager Product Development
Babcock & Wilcox
4674 Walena Drive
Medina, OH 44256
216/860-2674

Howard Hurwitz
Process Engineering
General Electric
5 Penn Plaza
New York, NY 10001
212/613-3175

Akitaka Ide
CT-121 Task Force Manager
Chiyoda Intern'l. Corp.
1300 Park Place Bldg.
Seattle, WA 98101
206/624-9350

Paul A. Ireland
Manager,Technical Staff-Power
Stearns Catalytic Corp.
P.O. Box 5888
Denver, CO 80217
303/692-3420

Yoshiro Ito
Engineering Manager
Mitsui Mining Company Ltd.
2-1-1, Nihonbashi-Muromachi, Chu-Ku
Tokyo 103 JAPAN
(03) 241-1339

Sam Jacobsson
Representative Manager
The Munters Corp.
P.O. Box 6428
Fort Myers, FL 33907
813/936-1555 X198

Thomas H. James
Project Manager Power
Betz Labs., Inc.
33 River Street
Chagrin Falls, OH 44022
N/A

James B. Jarvis
Staff Engineer
Radian Corporation
P.O. Box 9948
Austin, TX 78766
514/454-4797

Eva C. Jedruch
Manager Market Research
Dravo Wellman
4800 Grand Avenue
Pittsburgh, PA 15225
412/777-5138

Xiao-Lin Jiang
Student
University of Cincinnati
Dept. of Civil & Envrmntl. Engr.
Cincinnati, OH 45221
N/A

Howard J. Johnson
Air Quality Engineer
Ohio EPA
361 E. Broad Street
Columbus, OH 43216-1049
614/466-6116

Ken Johnson
Environmental Engineer
Duke Power Company
522 S. Church Street
Charlotte, NC 28242
704/373-5989

Lyndon Johnson
Air Quality Supervisor
St. Johns River Power Park
11201 N. Berlin Road
Jacksonville, FL 32226
904/766-5458

Robert M. Johnson
Group Mgr-Research
Gold Band Bldg. Products
1650 Military Road
Buffalo, NY 14217
716/873-9750

Charlie Johnston
Marketing Manager
Badische Corp.
602 Copper Road
Freeport, TX 77541
409/238-6327

Gary D. Jones
Senior Engineer
Radian Corporation
P.O. Box 9948
Austin, TX 78766
512/454-4797

Julian W. Jones
Senior Chemical Engineer
U.S. EPA
IERL, MD-61
Research Triangle Park, NC 27711
919/541-2489

Loy W. Jones
Salesman
Pfavdler Co.
3914 Miami
Cincinnati, OH 45227
513/561-8068

Stu Jones
Marketing Manager
MPSI
P.O. Box M312
York, PA 17402
717/843-8671

Claus Jorgensen
Chemical Engineer
Alurex Corp.
4915 Prospectus Dr.
Durham, NC 27713
919/544-4535

John A. Jorgensen
Vice President
Kissick Corp.
108 Benson East
Jenkintown, PA 19046
215/885-6650

Wojeiech Jozewicz
n/a
University of Texas
Dept. of Chemical Engineering
Austin, TX 78712-1165
512/471-7230

John F. Kamody
Project Manager
Koppers
Koppers Bldg.
Pittsburgh, PA 15219
412/227-2920

Dale Kanary
Envnmtl. & Special Proj. Engr.
Ohio Edison Company
76 S. Main Street
Akron, OH 44308
216/384-5744

Marilyn Kaplan
Editor
The McIlvaine Co.
2970 Maria Avenue
Northbrook, IL 60062
312/272-0010

Sumana Katchanat
Student
University of Cincinnati
Dept. of Civil & Envrmntl. Engr.
Cincinnati, OH 45221
N/A

Steve Katzberger
Supv., Emission Control
Sargent & Lundy
55 E. Monroe
Chicago, IL 60603
312/269-6672

N. Kawamura
n/a
Chiyoda Chemical Engr. & Const.
1300 Park Place Bldg.
Seattle, WA 98101
206/624-9350

Ronald H. Kaye
Executive Vice President
Peabody Process Systems, Inc.
201 Merritt 7
Norwalk, CT 06852
203/846-1600

Timothy C. Keener
Assistant Professor
University of Cincinnati
Dept. of Civil & Env. Eng.
ML #71, Cincinnati, OH 45221
513/475-5721

Donald J. Kelly
Engineer
Louisville Gas & Electric
5252 Cave Run Road
Louisville, KY 40216
502/566-4547

James R. Kendle
Regional Manager
FMC Corp. - AQC
1501 Woodfield Rd., Ste. 300
E. Schaumburg, IL 60195
312/843-1700

Donald Kerivan
Senior Engineer
Dravo Wellman Co.
4800 Grand Ave. Neville Is.
Pittsburgh, PA 15225
412/777-5470

Frank S. Kilvinger
President
Boyle Engineering Labs
207 North Woodwork Lane
Palatine, IL 60067
312/991-6200

Gary Kizior
Research Engineer
Amoco
N/A
Naperville, IL 60143
312/420-5999

Jonas Klingspor
Chemical Engineer
Battelle Columbus Labs
505 King Avenue
Columbus, OH 43212
N/A

W. D. Knowles
Senior Staff Engineer
Shell Oil Co.
P.O. Box 3105
Houston, TX 77001
713/241-1791

Bernard J. Koch
Director, Project Development
Conoco Coal Research Division
4000 Brownsville Rd.
Library, PA 15129
412/854-6612

G. H. Koch
Asso. Mgr. Corrosion Section
Battelle Columbus
505 King Avenue
Columbus, OH 43201
614/424-4430

Robert E. Kohl
Mgr. Appl. Engrng.
Environmental Elements Corp.
3700 Koppers Street
Baltimore, MD 21203
301/368-7375

Ken Kondo
Manager, FGD Systems
Mitsubishi Heavy Ind. America, Inc.
520 Madison Avenue
New York, NY 10022
212/826-2188

Pramod B. Koradia
Manager, Selexol Technology
Norton Co.
P.O. Box 350
Akron, OH 44309
216/688-4384

Jeffrey C. Korth
Sales manager
Effox, Inc.
9759 Inter Ocean Drive
Cincinnati, OH 45240
513/874-8915

Wes Kouns
Engineer
Armco Auto Makers
7077 Winchester
Boulder, CO 80301
303/530-1600

Hal Kruger
Engineer
Cleveland Electric Illuminating
P.O. Box 5000
Cleveland, OH 44101
216/622-9800 X3146

John H. Kuehl
Utility Sales Supervisor
Warman International, Inc.
2701 S. Stoughton Road
Madison, WI 53716
608/221-2261

Peter Kutemeyer
General Manager
Bischoff
135 Cumberland Road
Pittsburgh, PA 15237
412/364-8860

Jeff Lamb
Manager, Gas Technical Services
U.S. Steel - Limestone Operations
n/a
Rogers City, MI 49779
517/734-4765

Klaus Lambeck
Coal Researcher
Ohio Coal Development Office
65 East State Street
Columbus, OH 43215
614/466-1088

William Lane
Engineer
Bechtel Power Corp.
15740 Shady Grove Rd.
Gaithersburg, MD 20850
301/258-4452

Wes Langeland
Vice President
Martek, Inc.
85 Research Road
Hingham, MA 02043
617/749-6992

Ellen Lanum
Manager, Conf. & Travel
Electric Power Research Inst.
P.O. Box 10412
Palo Alto, CA 94303
415/855-2193

B. A. Laseke
Group Supervisor
PEI Associates, Inc.
11499 Chester Road
Cincinnati, OH 45246
513/782-4700

Dennis Laslo
Sr. Development Engineer
Peabody Process Systems
201 Merritt 7
Norwalk, CT 06851
203/846-1600

Dorothy P. Lau
Env. Studies Engineer
Ontario Hydro
700 University Ave.
Toronto, Ontario M5G 1X6
416/592-7306

Ronald Laws
Engineer
Tampa Electric
P.O. Box 111
Tampa, FL 33601
813/228-4111

Peter Lawson
Supv. Engr.-Envnmt. Engeering
Ontario Hydro
700 University Avenue
Toronto, Ontario M5G 1X6
416/592-5393

John Lawton
Chemical Engineer
TVA 312 MDB
Muscle Shoals, AL 35660
N/A
205/764-2721

George O. Layman
Dir. of Power Generation Resch.
Gulf Power Company
P.O. Box 1151
Pensocola, FL 32520
904/434-8354

Randy LaFollette
Assist. Supervisor
C.G.& E.
P.O. Box 107
Union, KY 41091
606/586-5600

Phil Leckner
Sr. Process Engineer
Davy McKee Corp.
4715 S. Florida Avenue
Lakeland, FL 33807-5000
813/646-7194

Dave Ledonne
Performance Engineer
Cin. Gas & Electric
P.O. Box 107
Union, KY 41091
606/586-5600

George C. Lee
Engineering Specialist
Bechtel Group, Inc.
50 Beale
San Francisco, CA 94105
415/768-4802

Marcel Lefevre
V.P., Director of Engineering
The Munters Corp.
P.O. Box 6428
Fort Myers, FL 33907
813/936-1555 X142

John Lehto
Senior Production Engineer
Northern States Power Co..
Rt. 1, Box 146
Becker, MI 55308
612/261-4100

Juergen H. Leimkuehler
Consultant Engineer
GEA Bochum
Pregelstrasse 5
D-4300 Essen, W. Germany
N/A

Charles C. Leivo
Manager, Market Development
EPD/Dresser Industries
601 Jefferson, #2728
Houston, TX 77002
713/750-2091

Richard W. Lemay
Supervisor Construction
Santee Corporation
One Riverwood Drive
Moncks Corner, SC 29461
803/761-4038

Seth Levine
Manager, Marketing Appl.
Ameribrom, Inc.
1250 Broadway
New York, NY 10001
212/244-1412

C. J. Lewis
Dir. Environmental Services
National Lime Assoc.
P.O. Box 15453
Lakewood, CO 80215
303/237-2948

William A. Liegois
Head, Industrial Process Dept.
Stanley Consultants, Inc.
Stanley Bldg., 225 Iowa Ave.
Muscatine, IA 52761
319/264-6457

Tom Lillestolen
Manager
Flakt, Inc.
9111 Cross Park Drive
Knoxville, TN 37901
615/693-7550

Victor Lin
Advisor
Industrial Technology Research Inst
1021 Kuang Fu Road
Hsinchu TAIWAN
N/A

Christina Lishawa
n/a
Southern Research Institute
P.O. Box 55305
Birmingham, AL 35255-5305
205/323-6592

Thomas M. Little
Chemical Engineer
Tennessee Valley Authority
P.O. Box 150
W. Paducah, KY 42086
502/443-9388

David Livengood
Section Manager
Argonne National Laboratory
9700 S. Cass Avenue
Argonne, IL 60439
312/972-3737

James R. Long
Vice President of Marketing
MEI Systems, Inc.
3121 W. Spencer
Appleton, WI 54914
414/725-7752

Robert L. Longfellow
Mgr., Marketing & Bus. Dev.
Dravo Wellman Co.
4800 Grand Ave., Neville Is.
Pittsburgh, PA 15225
412/777-5145

William Z. Looney
Project Planning Engineer
Southern Co. Srvices
P.O. Box 2625
Birmingham, AL 35202
205/877-7289

Phillip Lowe
Director Adv. Tech.
NVS Corp.
910 Clopper
Gaithersburg, MD 20854
301/258-2468

Philip S. Lowell
Consulting Chemical Engineer
P.S. Lowell & Co., Inc.
8868 Research Blvd, Ste 309
Austin, TX 78758
512/451-3513

David Lu
Chief of Industrial Programs
SWOAPCA
2400 Breekman St.
Cincinnati, OH 45214
513/251-8777

Clement H. Luken
Chemist
Emery Chemicals
4900 Este Avenue
Cincinnati, OH 45232
513/482-2384

Kim Lukens
Process Engineer
United Engineers
30th South 17th Street
Philadelphia, PA 19104
215/422-3812

R. R. Lunt
Project Engineering Manager
United Engineers & Constructors
30 So. 17th Street
Philadelphia, PA 19101
215/422-3595

Mario Luperi
Vice President, Corporate Sales
Wm. Steinen Mfg. Co.
29 E. Halsey Rd.
Parsippany, NJ 07054
201/887-6400

Mike Machusak
Maintenance Engineer
Arizona Public Service Co.
P.O. Box 188
Joseph City, AZ 86032
602/288-3381 X421

S. H. Magliente
Environmental Engineer
Tennessee Valley Authority
1150 Chestnut Street Tower II
Chattanooga, TN 37401
615/751-5654

Emmett Mahoney
Manager Heat Recovery
Combustion Engineering
Andover Road
Wellsville, NY 14895
716/593-2700

Jason Makansi
Associate Editor
Power Magazine
1221 Ave. of the Americas
New York, NY 10020
212/512-4239

John E. Makar
Regional Manager
FMC Corp. - AQC
1501 Woodfield Rd., Ste. 300
E. Schaumburg, IL 60195
312/843-1700

Gordon Maller
Staff Chemist
Radian Corp.
8501 Mo-Pac Blvd.
Austin, TX 78766
512/454-4797

Ghassem B. Manavizadeh
Chief Process Engineer
Research-Cottrell
P.O. Box 1500
Sommerville, NJ 08876
201/685-4000

K. N. Mani
n/a
Allied Corporate Technology
P.O. Box 1021R
Morristown, NJ 07960
201/455-5410

Norman Mansson
Sr. Regional Sales Mgr.
Koch Engineering
161 East 42nd
New York, NY 10017
212/682-5755

Joanna Markussen
Chemical Engineer
U.S. Dept. of Energy, P.E.T.C.
P.O. Box 10940
Pittsburgh, PA 15236
412/675-6170

Patrick Maroney
Project Manager
Brown & Caldwell
P.O. Box 8045
Walnut Creek, CA 94596
415/937-9010

Dale J. Marshall
Supervisor
Dayton Power & Light Co.
P.O. Box 468
Aberdeen, OH 45101
513/549-2641

James R. Martin
Vice President Operations
Combustion Engineering, Inc.
31 Inverness Ctr. Pkwy.
Birmingham, AL 35243
205/967-9100

Werner K. Mattusch
Project Engineering
Burns & Roe
800 Kinderkamack Road
Oradell, NJ 07649
201/265-2000 X2450

Jack R. Maurer
Market Development Mgr.
Allegheny Ludlum Steel
Research Center
Brackenridge, PA 15014
412/339-5083

Michael A. Maxwell
Chief, Emissions/Effluent Tech.
U.S. EPA
MD-61
Research Triangle Park, NC 27711
919/541-2578

William J. McCabe
Design Engineer II
Duke Power Co.
P.O. Box 33189
Charlotte, NC 28212
704/373-8764

J. L. McDonald
n/a
Louisville Gas & Elec. Co.
311 W. Chestnut Street
Louisville, KY 40232
502/566-4612

Marilyn McIlvaine
Editor
McIlvaine Co.
2970 Maria
Northbrook, IL 60062
312/272-0010

R. W. McIlvaine
President
The McIlvaine Company
2970 Maria Avenue
Northbrook, IL 60062
312/272-0010

M. L. McInnis
Mgr.-Property & Envntl. Affairs
Louisville Gas & Electric Co.
311 W. Chestnut St.
Louisville, KY 40232
502/566-4612

Robert B. McKeagney
Mgr. Business Development
Vulcan Materials Company
P.O. Box 7497
Birmingham, AL 35253
205/877-3134

John C. McKenna
Regional Sales Manager
Ershings, Inc.
3575 Spruce Street
New Waterford, OH 44445
216/457-7635

John D. McKenna
President
ETS, Inc.
3140 Chaparral Drive
Roanoke, VA 24018
703/774-8999

John C. McMillan
N/A
Kissick Corporation
10489 Cinderella Drive
Cincinnati, OH 45242
513/791-6583

Michael L. Meadows
Pollution Control Engineer
Black & Veatch
P.O. Box 8405
Kansas City, MO 64114
913/967-7425

Jack Mehok
Utilities Mgr.
American Maize
1100 Indianapolis Blvd.
Hammond, IN 46326
219-659-2000 X319

Rajiv R. Mehta
Process Devlop. Engr.
FMC Corporation
1501 Woodfield Rd., Ste 300 East
Schaumburg, IL 60195
312-843-1700

M. T. Melia
Project Manager
PEI Associates, Inc.
11499 Chester Road
Cincinnati, OH 45246
513/782-4700

R. T. Melloan
Mgr., Property & Envrmntl. Affairs
Louisville Gas & Elec. Co.
311 W. Chestnut Street
Louisville, KY 40232
502/566-4612

Francis Merlie
Engineer
Philadelphia Electric Co.
2301 Market Street
Philadelphia, PA 19101
215/874-8451

James A. Messersmith
Operations Superintendent
St. Johns River Power Park
11201 New Berlin Road
Jackson, FL 32226
904/766-5430

James E. Meyers
Chemical Engineer
Detroit Edison
3331 W. Big Beaver
Troy, MI 48084
313/649-7330

H. J. Michaels
Project manager
Koppers Co., Inc.
Pittsburgh, PA
N/A
412/227-2563

M. G. Milobowski
ES EQPT Design Engineer
Babcock & Wilcox
Van Buren
Barberton, OH 44203
216/860-6630

J. E. Mirabella
Project Engineer
Chiyoda Intern'l. Corp.
1300 Park Place Bldg.
Seattle, WA 98101
206/624-9350

Dan Mitas
Systems Engineer
Black & Veatch
P.O. Box 8405
Kansas City, MO 64114
913/967-2189

J. D. Mobley
Environmental Engineer
U.S. EPA
IERL, MD-61
Research Triangle Park, NC 27711
919/541-2350

James D. Moore
Product Specialist
Nalco Chemical Co.
2901 Butterfield Road
Oak Brook, IL 60516
312/887-7500 X1645

Tom Morasky
Program Manager
Electric Power Research Inst.
3412 Hillview Avenue
Palo Alto, CA 94303
415/855-2468

David S. Morey
Supv. Marketing Research
Allied Corpoporate Technology
1411 Broadway
New York, NY 10018
212/391-5149

C. Morris
Manager, Business Development
Sim-Chem Limited
P.O. Box 17 Cheadle Hulme
Cheadle, Cheshire SK8 5BR England
061-485 6131

Rob Moser
Project Manager
Electric Power Research Institute
P.O. Box 10412
Palo Alto, CA 94303
415/855-2000

John B. Moss
Manager
Georgia-Pacific
133 Peachtree St.
Atlanta, GA 30303
404/521-5668

A. W. Mueller
Manager, Process Development
Church & Wright Co., Inc.
20 Kingsbridge Road
Piscataway, NJ 08854
201/885-1220

Edward J. Muren
Regional Manager
Research-Cottrell
85 W. Algonquin Rd., Ste 500
Arlington Hts., IL 60103
312/228-6228

William Murray
Mgr. Environmental Quality
ARCO Petroleum Products Co.
400 E. Sibley
Harvey, IL 60426
312/210-3598

John Mycock
Vice President
ETS, Inc.
3140 Chaparrall Drive
Roanoke, VA 24018
703/774-8999

Don Mzyk
Operations Supervisor
Texas Utilities Generating Co.
Box 651
Tatum, TX 75691
214/836-2711 X367

Jim Nail
Chemical Engineer
Tampa Electric
P.O. Box 111
Tampa, FL 33601
813/248-8365

Peter A. Nassos
Sr. Chemist
Radian Corp.
8501 MoPac Blvd.
Austin, TX 78766
512/454-4797

L. G. Neal
President
Noxso Corporation
P.O. Box 586
Library, PA 15129
412/675-6293

James K. Neathery
Production Engineer
East KY Power
P.O. Box 707
Winchester, KY 40391
606/744-4812

Daniel E. Neighbor
Marketing Manager
Church & Dwight Co., Inc.
20 Kingsbridge Road
Piscataway, NJ 08854
201/-885-1220

Millard W. New
Plant Manager
Southern Indiana Gas & Elec.
20-24 NW 4th
Evansville, IN 47741
812/985-6208

Thomas H. Newhams
Vice President Engineering
Peabody Process Systems, Inc.
201 Merritt 7
Norwalk, CT 06852
203/846-1600

Linda G. Niman
Process Engineer
Pacific Power & Light
920 SW 6th
Portland, OR 97204
503/243-3021

Baker N. Nimry
Group Leader
Nalco Chemical Co.
1601 Diehl Road
Naperville, IL 60566
312/961-9500

Paul S. Nolan
Manager, SO2 Removal Programs
Babcock 7 Wilcox
20 So. Van Buren Avenue
Barberton, OH 44203
216/860-1074

Richard D. Norton
Regional Manager
Pathway Bellows
P.O. Box 880
Addison, TX 75001
214/380-8300

Robert D. O'Hara
Environmental Engineer
Duquesne Light Co.
301 Grant Street
Pittsburgh, PA 15279
412/393-6098

Ron Ojanpera
Gen. Sales Mgr.
Joy Canada
235 King Street East
Kitchener, Ontario N2G 4N6
519/578-6400

Alan Olderstein
n/a
Mixing Equipment Co., Inc.
135 Mt. Read Blvd.
Rochester, NY 14611
716/436-5550

Earl D. Oliver
Consultant
Pacific Gas & Electric
2049 Kent Drive
Los Altos, CA 94022
415/964-4838

James L. Olsen
Sales Engineer
Brand Industrial Serv., Inc.
1420 Renaissance Drive
Park Ridge, IL 60068
312/298-1200

David G. Olson
Manager, Utility Sales & Planning
GE Environmental Serv., Inc.
200 N. 7th Street
Lebanon, PA 17042
717-274-7355

Lars Olsson
Vice President, International
Flakt, Inc.
146-148 Sound Beach Ave.
Old Greenwich, CT 06870
203/637-5401

Sidney R. Orem
Executive Director
Industrial Gas Cleaning Inst.
Suite 304, 700 North Fairfax St.
Alexandria, VA 22314
703/836-0480

Delbert M. Ottmers, Jr.
Asst. Vice President
Radian Corp.
P.O. Box 9948
Austin, TX 78766
512/454-4797

Joseph H. Oxley
Project Manager
Battelle
505 King Avenue
Columbus, OH 43201
614/424-7888

Matt W. Packbier
Vice President
Hoechst-Uhde Corporation
560 Sylvan Avenue
Englewood Cliffs, NJ 07632
201/569-0100

Rashmi Parekh
Resident Sales Manager
Dorr Oliver, Inc.
274 Riverside Avenue
Westport, CT 06880
203/358-3802

Joe Parent
Operations Specialist
Florida Power & Light
11201 New Berlin Road
Jacksonville, FL 32226
904/766-5425

Del Parsons
Sales Manager
Bete Fog Nozzle, Inc.
324 Wells Street
Greenfield, MA 01301
413/772-0846

John S. Pasquale
Project Engineer
Peabody Process Systems, Inc.
201 Merritt 7
Norwalk, CT 06852
203/846-1600

Jayanti R. Patel
Engineer
Philadelphia Electric Co.
2301 Market Street
Philadelphia, PA 19101
215/841-5042

Henry W. Pennline
Chemical Engineer
U.S. Department of Energy
P.O. Box 10940
Pittsburgh, PA 15236
412/675-6013

Michael Perlsweig
Program manager
U.S. Dept. of Energy
FE-23
Washington, DC 20545
301/353-4399

Robert O. Petkus
Senior Staff Engineer
FMC Corp. - AQC
1501 Woodfield Rd.
E. Schaumburg, IL 60195
312/843-1700

Peter H. Phillips
Manager, I&C, FGD Systems
Wheelabrator Air Pollution Control
600 Grant Street
Pittsburgh, PA 15219
412/288-7355

Richard F. Pierce
Supv.-Acid Dept. Strategy
Wis. Dept. of Natural Resources
101 S. Webster Street
Madison, WI 53707
608/266-8470

Wallace Pitts
Associate
Killzelly Associates
P.O. Box 31265
Raleigh, NC 27627
919/781-3150

James F. Plappert
Vice President, Marketing
Conversion Systems, Inc.
115 Gibraltar Road
Horsham, PA 19044
215/441-5900

Everett L. Plyler
Director, Gas Cleaning Div.
U.S. EPA
MD-60
Research Triangle Park, NC 27711
919/541-2918

Franz G. Pohl
Research Physicist
Southern Research Institute
P.O. Box 55305
Birmingham, AL 35255
205/323-6592

Wade Ponder
Asst. to Lab Director
U.S. EPA/AEERL
MD-60
Research Triangle Park, NC 27711
919/541-2818

Thomas L. Potter
Exec. Director
National Lime Asso.
3601 N. Fairfax Drive
Arlington, VA 22201
N/A

Paul R. Predick
Project Manager
Sargent & Lundy
55 E. Monroe
Chicago, IL 60603
312/269-6671

Sheldon Presser
Supervisor, FGD Systems
Santee Cooper
P.O. Box 98
Cross, SC 29436
803/351-4586

Frank T. Princiotta
Dir., Air & Energy Engr. Research
U.S. EPA
IERL, MD-60
Research Triangle Park, NC 27711
919/541-2821

Richard Rao
Consultant
Ebasco Services
160 Chubb Avenue
Lyndhurst, NJ 07071
201/460-5903

Frederick I. Rapp
Sales Representative
Armod, Inc.
7000 Roberts
Kansas City, MO 64125
816/242-5452

Jay Ratafia-Brown
Fossil Energy Projects Dir.
Science Applications Int'l. Corp.
1710 Goodridge Drive
McLean, VA 22180
703/448-6343

Bebe Raupe
O.K.I. Correspondent
Bureau of National Affairs
3821 Fox Run
Cincinnati, OH 45236
513/793-1753

Steven F. Rees
Marketing Manager
Clean Air Engineering, Inc.
207 North Woodwork Lane
Palatine, IL 60067
312/991-3300

Bruce W. Remick
Director of Aglime Marketing
National Stone Associationa
1415 Elliot Place, NW
Washington, DC 20007
202/342-1100

Buddy C. Reynolds, Jr.
Market Research Analyst
Koppers Company, Inc.
Room 3000 Koppers Bldg.
Pittsburgh, PA 15219
412/227-2504

Richard Rhudy
Project Manager
Electric Power Research Inst.
P.O. Box 10412
Palo Alto, CA 94303
415/855-2421

Phillip Richardson
Development Manager
Dupont
1007 Market St.
Wilmington, DE 19898
302/774-4930

Ken Rieser
General Manager
Black River Lime Co.
P.O. Box 1
Butler, KY 41006
606/472-7721

Richard H. Riggs
Environmental Engineer
Tennessee Valley Authority
100 IBM Bldg.
Chattanooga, TN 37401
615/751-4420

J. D. Riley
Director, Pre-Compact Opers.
Lurgi Corporation
666 Kinderkamack Rd.
River Edge, NJ 07661
201/-967-4921

Ed Riordan
Supv. of Waste Treatment
SWLT
Municipal Bldg.
Springfield, IL 62701
217/789-2238

R. F. Robards
Chemical Engineer
Tennessee Valley Authority
1160 Chestnut Street Tower II
Chattanooga, TN 37401
615/751-5662

Marline A. Roberts
Scientist
Indianapolis Power & Light
P.O. Box 1595B
Indianapolis, IN 46206
317/261-8879

Gary T. Rochelle
Professor
University of Texas
Dept. of Chemical Engineering
Austin, TX 78712
512/471-3434

Byron C. Rodgers
Group Manager
Santee Cooper
One Riverwood Drive
Moncks Corner, SC 29461
803/761-4038

Frank Roma
Project Engineer
New York State Electric & Gas
4500 Vestal Pkwy.
Binghamton, NY 13902
607/729-2551 X4529

Jean T. Rose
Project Supervisor
Betz Laboratories, Inc.
4636 Somerton Road
Trevose, PA 19047
215/355-3300

Harvey S. Rosenberg
Senior Research Engineer
Battelle Columbus Labs.
505 King Ave.
Columbus, OH 43201
614/424-5010

Clifford A. Rosene
FGD Production Supervisor
Northern Indiana Pub. Serv.
Route 1, Box 320
Wheatfield, IN 46392
219/956-5180

R. W. Ross
Industry Manager
Inco Alloys
P.O. Box 1958
Huntington, WV 25720
304/526-5562

Edward S. Rubin
Director, CEES
Carnegie-Mellon University
Schenley Park
Pittsburgh, PA 15213
412/578-2491

Donald H. Rullman
Vice President
Lurgi Corporation
666 Kinderkamack Road
River Edge, NJ 07661
201/967-4914

Richard A. Runyan
Projecrs Manager
Tennessee Valley Authority
1160 Chestnut Street Towers Ii
Chattanooga, TN 37401
615/751-5663

R. E. Rush
Section Manager
Southern Co. Services
P.O. Box 2625
Birmingham, AL 35213
205/870-6320

Christi L. Rust
Marketing Services Coordinator
FMC Corp. - AQC
1501 Woodfield Rd., Ste. 300
Schaumburg, IL 60195
312/843-1700

Stefan J. Rzad
Staff
General Electric
P.O. Box 8
Schenectady, NY 12301
518/387-7185

A. Saleem
V.P., Marketing & Technology
G.E. Environmental Services, Inc.
200 N. 7th Street
Lebanon, PA 17042
717/274-7171

Joseph J. Saliga
Senior Mechanical Engineer
Fluor Engineers
200 W. Monroe
Chicago, IL 60130
312/368-3852

N. C. Samish
Staff Engineer
Shell Devel. Co.
Box 1380
Houston, TX 77001
713/493-7944

Michael A. Sandell
Manager
Wheelabrator Air Pollution Control
600 Grant St.
Pittsburgh, PA 15219
412/288-7519

Chakra J. Santhanam
Manager
Arthur D. Little, inc.
20 Acorn Park
Cambridte, MA 02140
617/864-5770 X5423

Harry Sargent
Project Leader
Dow Chemical
n/a
Freeport, TX 77541
409/238-5843

Otto M. Sarmiento
Sr. Development Engineer
Peabody Process Systems
201 Merritt 7
Norwalk, CT 06851-6037
203/846-1600

Alvin Saunders, Jr.
Supt. Technical Services
Santee Cooper
P.O. Box 1275
Georgetown, SC 29442
803/546-4171

Joseph Schival
Engineer
Philadelphia Electric Co.
RD 3 Box 12
Phoenixville, PA 19460
215/933-8995

Brian A. Schmidt
Asst. Chemical Engineer
Union Electric Co.
P.O. Box 149
St. Louis, MO 63166
314/554-3093

David Scott
Proposal Engineer
Babcock & Wilcox Co.
20 S. Van Buren Avenue
Barberton, OH 44203
216/860-2633

Ralph C. Searle
Mgr. Business Development
Coors Porcelain
600 9th Street
Golden, CO 80401
303/277-4368

David O. Seaward
Division Manager
The Munters Corp.
P.O. Box 6428
Fort Myers, FL 33907
813/936-1555 X132

Charles B. Sedman
Environmental Engineer
U.S. EPA
IERL, MD-61
Research Triangle Park, NC 27711
N/A

John M. Seibel
Supr. of FGD
Cin. Gas & Electric
Cincinnati, OH
N/A
606/586-5600

Nat Sekhar
Manager, Process Technology
GE Environmental Serv., Inc.
200 N. 7th Street
Lebanon, PA 17042
717/274-7195

M. Seminel
Dept. for Ind. Pollution Prevention
Agency for the Quality of the Air
n/a
n/a
N/A

Roland Seward
Process Engineer
Kennedy Van Saun
350 Railroad St.
Danville, PA 17821
717/275-6779

Raymond J. Shaffery
Director, Commercial Devel.
Church & Dwight Co., Inc.
20 Kingsbridge Road
Piscataway, NJ 08854
201/885-1220

Robert W. Shane
FGD Maintenance Supervisor
Indianapolis Power & Light Co.
RR #1
Petersburg, IN 47567
812/354-8801

Ike Shank
Marketing Sales Mgr.
Ashland Chemical
P.O. Box 2219
Columbus, OH 43216
614/889-3785

Paul S. Shapiro
Environmental Engineer
U.S. EPA
401 M St., S.W. RD-681
Washington, DC 20460
202/382-2583

Donald M. Shattuck
Senior Process Engineer
Stearns Catalytic Corp.
P.O. Box 5888
Denver, CO 80111
303/692-4139

R. M. Sherwin
Poll. Control Supvr.
Bechtel Group, Inc.
P.O. Box 3965
San Francisco, CA 94119
415/768-6135

Atul C. Sheth
Associate Professor
Univ. of Tennessee Space Inst.
Tullahoma, TN 37388
N/A
615/455-0631

Yei-Shong Shieh
Director, Technical Services
Conversion Systems, Inc.
115 Gibraltar Road
Horsham, PA 19044
215/441-5900

Joe Shortt
Tech. Sales Rep.
USG Industries
P.O. Box 417
Saltville, VA 24370
703-496-4690

Mark S. Siegler
Chief Technical Supporter
EPA Washington DC
401 M. Street SW
Washington, DC 20460
202/382-2835

William Silence
Engineering Associate
Cabot Corporation
1020 West Park Avenue
Kokomo, IN 46902-9013
317/456-6201

A. P. Simko
Manager, Technical Services
Arizona Public Service Co.
P.O. Box 21777, Station 5760
Phoenix, AZ 85036
602/271-7261

Jerry L. Simpson
Boiler Engineer
Central Illinois Public Serv. Co.
607 East Adams Street
Springfield, IL 62701
217/523-3600

Dave G. Skamenca
Industry Sales Manager
Komline-Sanderson
12 Holland Avenue
Peapack, NJ 07977
201/234-1000

A. V. Slack
Consultant
SAS Corp.
Wilson Lake Shores
Sheffield, AL 35660
205/383-1627

Tommy J. Slater
FGD Supervisor
Southwestern Electric Power Co.
Rt. 2, Box 165
Hallsville, TX 75650
214/935-3181

Alfred A. Slowik
Environmental Affairs
Pennsulvania Electric Co.
1001 Broad St.
Johnstown, PA 15907
814/533-8217

Michael Smart
Dir. Prod. Devl. Tech. Serv.
J.P. Stevens & Co., Inc.
P.o. Box 208
Greenville, SC 29602
803/239-4830

David E. Smith
Project Manager
FMC Corporation
2000 Market Street
Philadelphia, PA 19103
215/299-6806

David L. Smith
Research Engineer
Pfizer, Inc.
640 N. 13th Street
Easton, PA 180423
215/250-3374

Ken Smith
Senior Process Engr. Specialist
Environment Ontario
880 Bay St., 4th Floor
Toronto, Ontario CANADA
416/965-5776

Morton J. Smith
Director,Plant Improvement Programs
Pickarad,Lowe and Garrick, Inc.
2260 University Avenue
Newport Bech, CA 92060
714/650-8000

Pete V. Smith
Group Supervisor
Babcock & Wilcox
1562 Beeson
Alliance, OH 44720
N/A

John V. Smolenski
FGD Specialist
Stone & Webster Engineering
3 Executive Campus
Cherry Hill, NJ 08034
609/482-3584

Donald R. Snider
Supr. of Technical Serv.
Cin. Gas & Electric
East Bend Station
Union, KY
606/586-5600

William W. Sommer
Sales Engineer
Stebbins Engineering
363 Eastern Blvd.
Watertown, NY 13601
315/782-3000

Leslie E. Sparks
Senior Chemical Engineer
U.S. EPA
MD-61
Research Triangle Park, NC 27711
919/541-2458

J. A. Spegele
Marketing Manager-Lime
U.S.G. Industries
3959 Electric Road
Roanoke, VA 24014
703/732-3159

H. W. Spencer
Manager, Advanced Technology
Joy Manufacturing Co.
4565 Colorado Blvd.
Los Angeles, CA 90039
818/240-2300 X426

Jeff Stallings
Sr. Engineering Economist
SRI, International
333 Ravenswood Avenue
Menlo Park, CA 94025
415/859-3077

John Stange
Manager of Mechanical Engr.
Gibbs & Hill, Inc.
11 Penn Plaza
New York, NY 10001
212/760-4162

Larry Stanislow
Power & Food Industry Mgr.
Pennwalt Corp.
Three Parkway
Philadelphia, PA 19102
215/587-7185

Glenn G. Stauffer
Project Engineer
Pennsylvania Power & Light
2 No. 9th Street
Allentown, PA 18101
215/770-6563

George C. Stegmann
Principal Research Engineer
Con Edison Co. of NY
4 Irving Place, Rm. 1428
New York, NY 10003
212/460-2754

Howard Steiman
Senior Chemical Engineer
Boston Edison
800 Boylston St. M/C P276
Boston, MA 02199
617/424-3298

Sonya Stelmack
Chemical Engineer
U.S. EPA
401 M. Street
Washington, DC 20460
202/382-2851

W. K. Stemple
Manager Technical Contracting
Allied Corp.
P.O. Box 1021 R.
Morristown, NJ 07960
201/455-4571

Edward W. Stenby
Manager, Fossil Engineering
Stearns Catalytic Corp.
P.O. Box 5888
Denver, CO 80217
303/692-4134

John G. Stensland
Marketing Manager
FMC Corp. - AQC
1501 Woodfield Rd., Ste. 300
E. Schaumburg, IL 60195
312/843-1700

Carl Stevens
New Products Engineer
Lukens Steel Co.
Arc Bldg. B200
Coatesville, PA 19320
215/383-3208

Dorothy A. Stewart
Project Manager
Electric Power Research Inst.
P.O. Box 10412
Palo Alto, CA 94303
415/855-2609

Jack Stewart
Utility Marketing Manager
Babcock & Wilcox
20 S. Van Buren Avenue
Barberton, OH 44203
216/753-4511

Joseph G. Stites
Consultant
Consultant
203 Barley Mill Rd.
Old Hickory, TN 37138
N/A

Robert Storms
Division Manager
Stebbins Engineering
363 Eastern Blvd.
Watertown, NY 13601
315/782-3000

Don Stowe
Vice President
Dravo Lime Co.
3600 Neville Road
Pittsburgh, PA 15225
412/777-5574

Ann-Mari Stromberg
Research Engineer
Lund Inst. of Technology
P.O. Box 124
S-22242 Lund,Sweden
46 46 108277

Douglas S. Stroud
Mechanical Engineer
Tennessee Eastman Co.
P.o. Box 511
Kingsport, TN 37662
615/229-4299

Frederick A. Sudhoff
Chemical Engineer
Tennessee Valley Authority
501 CEB
Muscle Shoals, AL 35661
205/386-2514

Larry Sundberg
Marketing Manager
B. F. Goodrich
500 S. Main Street
Akron, OH 44318
216/374-3137

Carl E. Swanson
Sr. Metallurgical Engr.
Newmont Exploration Ltd.
44 Briar Ridge Road
Danbury, CT 06810
203/743-6784

Gregory Swanson
Scientist
S-Cubed, Div. of Maxwell Labs
P.O. Box 1620
LaJolla, CA 92038
619/453-0060 X467

Don Syler
Engineering Section Hd.
Consumers Power Co.
1945 Pavnull Rd.
Jackson, MI 49201
517/788-1946

Hank Taylor
Chemical Engineering Supv.
Public Service Co. of NM
P.O. Box 227
Waterflow, NM 87421
505/598-7834

Rod Thacker
Marketing Representative
Tenneco Minerals
3707 FM 1960 West St., Ste. 420
Houston, TX 77068
713/580-0847

Mack A. Thompson
Planning Engineer
Illinois Power Company
500 S. 27th Street
Decatur, IL 62525
217/424-6552

Edward W. Tomeo
Generation Engineer
Northeast Utilities
P.O. Box 270
Hartford, CT 06141
203/665-3791

Joseph M. Towarnicky
Staff Scientist
United McGill
P.O. Box 820
Columbus, OH 43216
614/443-0192

Paul E. Traccarella
Proposal Manager
Combustion Engineering
31 Inverness Center Pkwy.
Birmingham, AL 35243
205/967-9100

E. C. Trexler
Program Manager
U.S. DOE
Washington, DC 20545
N/A
301/353-2083

Ronald J. Triscori
Director of Products
Western Precipitation Div.
4565 Colorado Blvd.
Los Angeles, CA 90039
818/240-2300 X527

Dennis Tulenko
Development Engineer
Wheelabrator APC
600 Grant Street
Pittsburgh, PA 15219
412/288-7589

Lydon Turner
Process Engineer
Conoco, Inc.
P.O. Box 1267
Ponca City, OK 74603
405/767-3131

Jim Turton
Regional Sales Manager
ACDC, Inc.
1262 U.S. Rt. 50
Milford, OH 45150
513/248-1820

John P. Valentine
General Manager
Allied Corp.
7 Powder Horn Drive
Mt. Betel, NJ 07060
201/560-1750

Vino Verma
Sr. Engineer Technologist
Santee Cooper
One Riverwood Drive
Moncks Corner, SC 29461
803/761-4038

J. N. Voyles
Mgr., Property & Envrmntl. Affairs
Louisville Gas & Elec. Co.
311 W. Chestnut Street
Louisville, KY 40232
502/566-4612

Dave Wagner
Sales Representative
Marblehead Lime
300 W. Washington
Chicago, IL 60606
312/263-4490

John R. Walenten
Process Development Engineer
Badische Corp.
50 Central Ave.
Kearny, NJ 07032
201/589-1600

Bill Walker
Project Engineer
Bete Fog Nozzle, Inc.
324 Wells Street
Greenvield, MA 01301
413/772-0846

Carlton J. Walker, Jr.
n/a
Arthur D. Little, Inc.
15 Acorn Park
Cambridge, MA 02140
617/864-5770 X2689

Dan Walker
Technical Director
Genstar Lime Co.
2551 N. Green Valley Pkwy.
Henderson, NV 89015
702/451-5441

Richard G. Walker
Dir. of Mining & Envrn. Services
Gold Bond Building Products
2001 Rexford Road
Charlotte, NC 28211
704/365-0950

Arden Walters
Tech. Spec.
Florida Power & Light
P.O. Box 14000
Juno Beach, FL 33408
305/744-0369

K. H. Wang
Senior Development Engineer
FMC Corp. - AQC
1501 Woodfield Rd., Ste. 300
E. Schaumburg, IL 60195
312/843-1700

Dale R. Warner
National Sales Manager
Air Clean Damper Co.
P.O. Box 767428
Roswell, GA 30076
404/587-5293

Yoshiyuki Watanabe
Senior Engineer
Mitsubishi Heavy Industries, Ltd.
15-1, Tomihisa-cho, Ichigaya
Shinjuku-ku, Tokyo, Japan 162
TOKYO 03 355-7443

Greg F. Weber
Research Supervisor
UND Energy Research Center
15 North 23 Street
Grand Forks, ND 58202
701/777-5222

Chris Wedig
Scrubber Consultant
Stone & Webster
245 Summer St.
Boston, MA 02107
617/589-5737

Carl Weilert
Process Engineer
Burns & McDonnell Engineering
P.O. Box 173
Kansas City, MO 64141
816/333-4375

Joe Weinberg
Product Manager
IDE
1250 Broadway
New York NY 10001
212/244-1412

Andrew Weiss
Vice President
Stebbins Engineering
363 Eastern Blvd.
Watertown, NY 13601
315/782-3000

William Wellock
Asst. Power/Food Mgr.
Pennwalt Corp.
Three Parkway
Philadelphia, PA 19102
215/587-7185

William L. Wells
Director
Cnt. for Research on Sulfur in Coal
615 East Peabody Drive
Champaign, IL 61820
217/333-9241

James Welshimer
Manager, Technical Services
National Lime & Stone Co.
P.O. Box 120
Findlay, OH 45839
419/422-4341

Woodrow West
Marketing Mgr.
Babcock & Wilcox
350 S. Van Buren
Barberton, OH 44203
216/860-1013

Charles S. White
Sales Manager
Conversion systems, Inc.
115 Gibraltar Road
Horsham, PA 19044
215/441-5900

Jack White
Plant Chemist
New York State Elec. & Gas
7725 Lake Road
Parker, NY 14012
716/795-9501 X5050

Malcolm L. White
Research Scientist
Lehigh University
F. MacDonald Sinclair Memorial Lab
Bethlehem, PA 18015
215/861-3576

Bruce M. Whitmyre
Regional Representative
Wallace & Tiernan Div. Pennwalt
1900 Dublin-Granville Road
Columbus, OH 43229
614/895-0911

James H. Wilhelm
Consultant
Codan Associates, Inc.
2394 Charros Road
Sandy, UT 84092
801/571-6974

D. A. Williams
Sales Manager
Davy McKee
4715 S. Florida Avenue
Lakeland, FL 33803
813/646-7236

John E. Williams
Program Manager-Flue Gas Cleanup
U.S. Dept. of Energy/PETC
P.o. Box 10940C
Pittsburgh, PA 15236
412/675-5727

Frederick A. Willits
Mechanical Engineer
Dayton Power & Light
P. O. Box 1247
Dayton, OH 45401
513/224-6376

Leon E. Winget
Coal Specialist
Ohio Coal Development Office
65 East State Street
Columbus, OH 43215
614/466-1088

Mike Wisdom
Chemical Engineer
Tippett & Gee, Inc.
502 N. Willis
Abilene, TX 79603
915/673-8291

Tom Wood
Business Development Representative
Bechtel Group, Inc.
50 Beale
San Francisco, CA 94119
415/768-7489

Gary Woodall
Maint. Superintendent
East Kentucky Power
P.O. Box 398
Maysville, KY 41056
606/883-3165

Martin Wright
Product Sales Manager
FMC Corporation - ACD
2000 Market Street
Philadelphia, PA 19103
215/299-6818

Beth A. Wrobel
Engineer
Northern Ind. Public Serv. Co.
Rt. #1, Box 320
Wheatfield, IN 46392
219/-956-5251

S. C. Wu
Project Manager
Industrial Technology Research Inst
1021 Kueng Fu Road
Hsinchu TAIWAN
N/A

James E. Wuchter
Project Leader
Badische Corp.
602 Copper Road
Freeport, TX 77541
409/238-6284

Richard A. Yannuzzi
Great Lakes Reg. Mgr.
LTV Fibercast
P.O. Box 1869
Highland, IN 46322
312/964-6575

Joseph Yarze
Consultant
Allied Corp.
P.O. Box 1021R
Morristown, NJ 07960
201/560-1750

Kurt E. Yeager
V. P., Coal Combustion Sys. Div.
Electric Power Research Inst.
3412 Hillview Ave.
Palo Alto, CA 94304
415/855-2456

James T. Yeh
Chemical Engineer
U.S. Dept. of Energy
P.O. Box 10900
Pittsburgh, PA 15236
412/675-5737

Heeyoung Yoon
Research Group Leader
Conoco Coal Research Division
4000 Brownsville Road
Library, PA 15129
412/854-6662

Charles S. Young
Product Manager
Astro metallurgical
P.O. Drawer 520
Wooster, OH 44691
216/264-8639

Dean Young
President
Kissick Corp.
108 Benson East
Jenkintown, PA 19046
215/855-6650

Richard W. Young
District Manager
Combusion Engrg. Inc.
6840 Ashfield Drive
Cincinnati, OH 45242
513/984-6920

Kent D. Zammit
Design Mech. Engr.
L. A. Dept. of Water & Power
111 N. Hope St., Rm. 667
Los Angeles, CA 90051
213/481-5019

Gerhard Zeiler
President
Ekato Corp.
7 Pearl Ct.
Allendale, NJ 07401
201/825-4684

Jan T. Zmuda
Applications Engineer
Research Cottrell
P.O. Box 1500
Somerville, NJ 08876
201/685-4915